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"A FEEDFORWARD CONTROL SYSTEM FOR A DOUBLE EFFECT EVAPORATOR"

©

by

ALAN H. WILSON

A THESIS
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The undersigned certify that they have read and recommend to the Faculty of Graduate Studies for acceptance, a thesis entitled "A Feedforward Control System for a Double Effect Evaporator," submitted by Alan H. Wilson, in partial fulfillment of the requirements for the degree of Master of Science.

ABSTRACT

A feedforward concentration controller for a double effect evaporator has been developed on the basis of an overall, steady state material balance around an evaporating unit. The controller calculates the steam flow required by the evaporator using the system inputs feed flow rate, feed concentration and product concentration set point. The model used by the controller was designed to provide suitable dynamic compensation to the steam flow following an upset in any of these inputs. Feed flow upsets and product concentration set point changes were followed by immediate adjustment of the steam flow rate to its new steady state value. Following feed concentration disturbances, the steam flow rate was reset to approach its new steady state value via a first order response that matched the concentration response in the first effect. Three parameter settings representing the total heat losses from the evaporator, the feed latent heat requirements, and the reciprocal of the evaporator steam economy are also required by the controller for the computation of the steam flow requirements.

The controller was constructed from individual analog computing modules and integrated with the conventional, Foxboro electronic instrumentation on the unit operations laboratory double effect evaporator. This integration was made in such a manner that the evaporator could be run with no concentration control (open loop), with feedback or feedforward concentration control only, or with feedback trimmed, feedforward concentration control. The major evaporator control objective for all control systems, was to achieve close product concentration control

without using either feed concentration or feed flow rate as a manipulated variable.

A series of eighteen experimental runs was made in which disturbances in feed flow rate, feed concentration and product concentration set point were introduced to the evaporator system while it was under control of each of the alternate control systems mentioned above. Material and heat balance calculations were performed on the experimental data obtained at the initial and final steady state conditions during each run. Errors in these balances were generally very small: closure errors of less than five percent of the system heat or mass input were normal. The transient responses from each run were then qualitatively compared to yield an indication of the suitability of each type of product concentration control.

It was observed that feedforward concentration control alone did not always succeed in returning the product concentration to the set point following a disturbance, but that the transient period was shorter and less extreme than that of similar open loop runs. The product concentration response following disturbances to feed flow or concentration was smooth and quite rapid. The response following a concentration set point change was essentially the open loop concentration response to a step change in steam flow, and was slow.

The addition of negative feedback trimming to the controller output resulted in improved product concentration control after feed disturbances. The slight concentration oscillations which resulted were smaller than those which Andre[1] reported when feedback

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[1] H. Andre, A Mathematical Model for a Two-Stage Concentrating Evaporator, Doctoral Thesis, University of Alberta, Nov. 1966.

control alone was used to compensate for feed disturbances, and the transient period was also shortened. The feedforward-feedback control system did result in an improved product concentration response following feed disturbances, but its response to product concentration set point changes was very slow.

The experimental runs involving feedback concentration control only were made with the same major controller settings as presented by Andre. The resulting concentration transients were highly oscillatory and did not reach steady state during periods in excess of twice the two hour transient period he reported. These observations indicated that the results of this study were not directly comparable to those of Andre's previous study, because of the physical changes which were made to the evaporator during the interim. The most important changes and their consequences were:

1. A new circulation pump and large diameter line were installed in the forced circulation evaporator to increase the circulation rate through the second effect. The heat transfer coefficient was thereby noticeably increased and pump cavitation and liquid surging problems were largely eliminated.
2. The previous condenser condensate level control loop was replaced by a spillover level control system which eliminated condenser flooding.
3. External level indicators were installed on the first effect and second effect vapor-liquid separator. The level signals, which were transmitted from these indicators to their respective level controllers, gave noise free representation of the liquid levels in each effect.

4. A calibration tank unit was built to collect condensate from the first effect steam chest and the condenser. This allowed every stream leaving the evaporator to be collected separately for volumetric calibration.

5. The condenser vacuum control valve was removed from the vacuum line and relocated in an atmospheric bleed to the vacuum line. This pressure control system was completely automatic and provided excellent condenser vacuum control.

In addition to improving the evaporator performance via modifications to evaporator equipment and instrumentation, this study presents and justifies a feedforward control system for a double effect evaporator. It also points out the need for including at least the major process dynamics within the feedforward control system, and establishes a basis from which further investigation of evaporator control systems may be initiated once the evaporator is operating under the new Chemical and Petroleum Engineering data acquisition and control computer.

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CHAPTER I

INTRODUCTION

The design and modelling of a double effect evaporator was the topic of an earlier study by Andre (2). This evaporator was built in the unit operations laboratory and is the experimental equipment with which this study is also concerned.

The first effect is a short-tube-vertical or calandria-type evaporator; the second is a long-tube-vertical, forced circulation evaporator. The effects are connected in series for multiple effect operation. The sugar-water feed solution can be fed to either effect, although all studies to date have been on the feedforward system in which the vapor and solution flows through the unit are in the same direction.

Low pressure steam supplied to the first effect steam chest provides the energy required by the evaporation process, and a steam jet ejector supplies vacuum to maintain the second effect pressure (and therefore boiling temperature) well below that of the first effect. This pressure differential permits the steam produced in the first effect to condense in the steam chest of the second effect. The solution is thereby boiled a second time and its sugar concentration is increased in both effects--even though fresh steam is added only at the first effect. Since the laboratory evaporator consists of only two effects, the vapor produced in the second effect is condensed in a condenser and removed to tankage. The liquid discharge rate from each effect is determined by the respective liquid level controllers.

Four evaporator control schemes for this evaporator were tested and discussed by Andre [(2), pp. 61-68], but only one was successful in providing stable control of the product solution concentration. The three unstable concentration control systems were not able to cope with the ten minute time constants which characterize the liquid holdups in each effect. The successful system involved the control of product concentration via regulation of the steam flow to the first effect. The dynamics linking these two variables are primarily in the vapor phase and are quite rapid for the laboratory evaporator; therefore, this control system was able to compensate quickly for input disturbances--once they were detectable in the product concentration--and provided stable concentration control.

It appeared that the evaporator control system might be improved by letting a feedforward concentration controller determine the steam input flow rate. If all the measured variables required by this controller were either system inputs or set points, input disturbances could be detected before they actually entered the process. If the controller model could then be tuned to describe the process perfectly, feedforward steam flow compensation might succeed in preventing the disturbances from ever affecting the product concentration.

The main objective of this work was therefore to design, build and test a feedforward product concentration controller on the evaporating unit. Neither feed flow rate nor feed concentration was to be available as a manipulated variable for use by the feedforward controller or any of the Foxboro instrumentation on the evaporator. It was also hoped that some of the operational difficulties reported by

Andre [(2), p. 79] and observed during early trial runs could be overcome. Briefly, the objectives of this project can be placed into two classifications:

1. Improvement of evaporator instrumentation.
    (a) Design and install a feedforward concentration controller.
    (b) Check the performance of the controller by a qualitative comparison of the transient response data obtained using the following concentration control systems;
        (i) open loop,
        (ii) feedback control as used by Andre,
        (iii) feedforward control,
        (iv) a workable combination of feedforward and feedback control.

    The evaporator transient responses under each control system must be obtained following separate disturbances to the feed flow rate, feed concentration and product concentration set point.
2. Improvement of evaporator operation.
    (a) Increase the recirculation rate in the second effect in an attempt to improve the heat transfer coefficient in the forced circulation evaporator.
    (b) Improve the condensate level control system in the condenser to prevent flooding of the condenser at low operating pressures.
    (c) Improve the ease of calibrating flow transmitters on the evaporator.
    (d) Design a condensate collection system which will permit the condensed vapor streams from each effect to be collected separately during experimental runs.

(e) Reduce the noise content of many of the process variables.

The evaporator improvements outlined above are collectively part of a program of preparation for the switchover to direct digital control for the evaporator and some of the other unit operations laboratory equipment. It was hoped that the first classification of changes would indicate the type of control system best suited for the evaporator, and also illustrate some of the problems involved in building a feed-forward controller and interfacing it with a conventional electronic control system. The remaining changes were concerned with the development of a dependable evaporating system, from which accurate experimental data could be extracted with a minimum amount of running time and operator effort.

CHAPTER II

THEORY

Evaporation is fundamentally a heat transfer process involving the exchange of heat from an energy source, such as steam, to a boiling solution. As vapor is boiled away from the solution, the concentration of solute in the solution is increased. The overall efficiency of this separation process depends upon the effectiveness of the evaporator in transferring heat to the boiling solution and upon the ability of the system to separate the vapor boilup from the solution. Single stage evaporators may be generally classified as being either the forced or the natural circulation type. Some of the many subdivisions of these two classifications are discussed and illustrated extensively in the literature (5, 6, 11, 15, 19) and do not require further explanation in this chapter.

A. Multiple Effect Operation

It is common practice to operate two or more evaporating units, or effects, in a series fashion, with the overhead vapor from the first effect condensing on the heating tubes of the second effect and so on until the last effect is reached. This multiple effect principle requires only that the pressure (and therefore the boiling temperature) be reduced in each succeeding effect so that a temperature difference can be established in each effect. The total system temperature difference (the first effect supply steam temperature minus the last effect vapor temperature) is distributed automatically over the various effects of a multiple effect system in a manner which is determined by the relative ability of each effect to transfer heat. To place a number on this "relative ability

to transfer heat" an overall heat transfer coefficient is defined at each heat transfer surface. This coefficient gives the heat transfer rate per unit of heat transfer area, per degree of temperature difference for a unit of time, and is inversely proportional to the heat transfer resistance inherent in each effect.

The evaporating capacity of a multiple effect system (all effects being identical) can not be increased above that which would result if the total system temperature difference was applied across a single effect of that evaporator. However, the multiple effect system is able to reuse the steam produced in each effect in the steam chest of the next effect. With ideal solutions and no inefficiencies in heat transfer, the total weight of steam removed from the solution, per pound of supply steam, would be numerically equal to the number of effects. This ratio of total boilup to the steam supply rate is defined as the steam economy. Factors which reduce the steam economy from its ideal value are:

1. Convection and radiation heat losses.
2. Latent heat requirements of the feed when it does not enter the evaporator at its boiling point.
3. The increase in the heat of vaporization of the solvent from effect to effect as the boiling pressure is lowered.
4. Loss of temperature difference due to boiling point rise (BPR) of the solution as it becomes more concentrated in solute. For example a 45% sugar-water solution has a BPR of 2°F when it is boiled at 160°F. The steam produced above this solution will have 2° of superheat which must be lost before the heat of vaporization can be obtained by condensing this steam.

The resistance to heat transfer in each effect of a multiple effect unit is determined by the evaporator design. If inert gases are continuously removed from the vapor space of each effect by a suitable purge system, there is essentially a single phase in the vapor space of each effect. With the evaporation rate in the first effect and the cooling water flow rate to the condenser on the last effect constant, the system heat transfer resistances will then determine the steady state temperature and pressure profiles within the evaporator.

When any one evaporating effect requires too large a portion of the total temperature difference and/or pressure drop in order to maintain the flow of heat through the system, it is an indication that the heat transfer resistance of that effect is too high. In the extreme case, the small capacity of one effect may prevent one or more other effects from operating at their design capacities. If the resistance of this effect cannot be decreased, it is purely a question of economics which must determine the manner in which this effect is used, or whether it should be used at all. The problem of operating evaporators of different capacities as a multiple effect unit will not be considered in this report because it is basically a problem in optimization; not in systems control.

A common method by which multiple effect evaporators may be classified is by the direction of the solution flow through the unit. The solution flow is in the same direction as the vapor flow for a forward feed system. In a backward, reverse, or counterflow system, the dilute feed enters the last effect and the product is withdrawn from the first effect. With a parallel feed system, the dilute solution is fed simultaneously to all effects. These feed systems may also be combined to form

a mixed feed system. For example, it is common in the paper industry for black liquor evaporators to be operated on a backward feed basis with the last two effects operated in parallel at the same pressure.

In backward feed systems, the most concentrated solution is boiled in the first effect where the temperature is highest and the product viscosity is minimal. This is desirable because the coefficient of heat transfer to a boiling solution is decreased as the solution viscosity increases. A backward feed system thereby maintains the highest possible evaporating capacity for a given amount of heat transfer surface, and is used to advantage when a viscous but non-heat sensitive solution must be concentrated. It has the disadvantage of requiring inter-effect pumps to transfer the solution from each effect to the preceeding effect. Forward feed systems are used when the product is heat sensitive and must be produced at a minimum temperature, or when it is desirable to eliminate inter-effect pumps. Sometimes two evaporator bodies are operated with parallel steam and feed streams so that the effective heat transfer area of one effect can be increased. This practise increases the evaporator capacity but not its steam economy.

B. Selection of Multiple Effect Evaporator Control Systems

In attempting to design or evaluate a control system for a process such as a multiple effect evaporator, it is essential that the interactions and steady state relationships which exist among many of the process variables be kept clearly in mind. When this is done, it is evident that the only difference between the control instrumentation for single and multiple effect evaporators lies in the increased number of level controllers which the latter will require. The multiple effect

evaporator is not however as easy to control as a single effect, due to the complications which the series connection of individual time constants adds to the control system.

The major controllable variables in the evaporation process are:

1. feed flow rate,
2. feed concentration,
3. solution liquid levels,
4. product concentration,
5. (a) supply steam temperature or pressure or (b) steam flow rate,
6. temperature or pressure of the vapor produced in the last effect.

In designing the control system it therefore becomes necessary to specify which of the process variables are to be controlled, manipulated, or left floating. This selection must be made in such a manner that:

1. the evaporator objectives with respect to product and material balance control are satisfied,
2. the system constraints are not violated,
3. the control system will not become unstable following a reasonable disturbance to any of the input or load variables.

C. Steady State Steam Requirements for a Multiple Effect Evaporator

To design a feedforward controller for an evaporator it is necessary to develop a model of the unit which can be used to describe the major controlled variable in terms of the system input variables. The simplest, and perhaps most obvious, model which can be used to compute the steam flow that will be required to concentrate the feed solution to a specified value, is a steady state material balance. The basic process may be considered simply as a separation of a feed stream W_o into a

product stream W_p and a condensed vapor stream W_e. This separation requires the addition of energy to the process in the form of a steam flow of S pounds per minute. Figure 1 is a schematic diagram of this process.

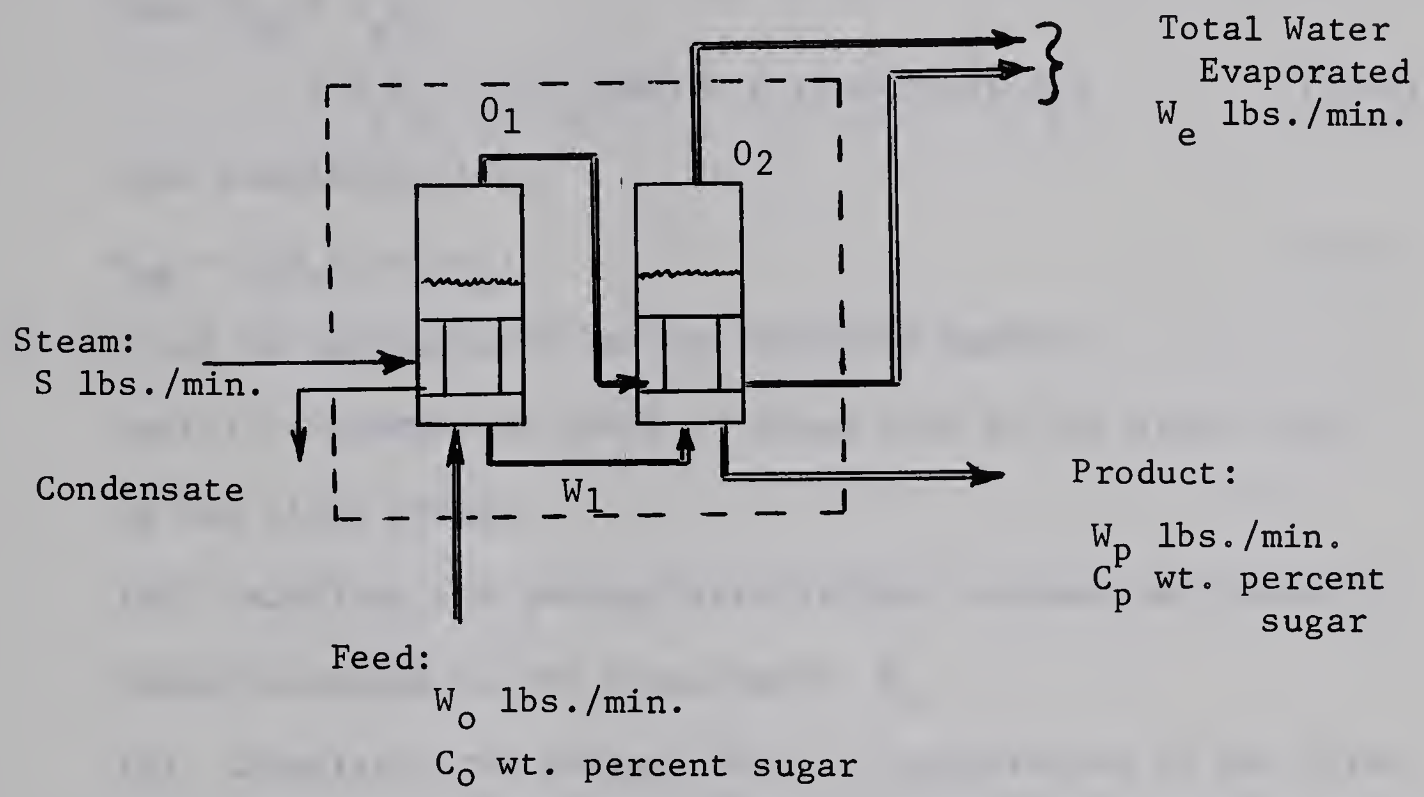


Figure 1 SCHEMATIC REPRESENTATION OF AN EVAPORATION PROCESS

1. The water content of the product stream can be determined at steady state by mass balances on the water and sugar flows:

   Water content of feed = $W_o\ (1-C_o/100)$

   Water content of product = $W_p\ (1-C_p/100)$ (II-1)

   Sugar input $W_oC_o/100$ equals the sugar output $W_pC_p/100$.

   Therefore $W_p = W_oC_o/C_p$ (II-2)

   By substituting (II-2) in (II-1) the water content of the product is $\frac{W_o\ C_o(1-C_p/100)}{C_p}$ (II-3)

The water which must be evaporated from the system is therefore

$$W_e = W_o(1-C_o/100) - W_oC_o(1-C_p/100)/C_p \qquad \text{(II-4)}$$

2. The steam flow requirement can now be determined if K_s is defined as the ratio of the flow of supply steam to the total rate of

evaporation.

i.e. $K_s = \dfrac{\text{lb. steam required}}{\text{lb. water evaporated}}$ (II-5)

Thus $S_{MB} = K_s W_e$

$= K_s W_o [\,(1-C_o/100) - C_o(1-C_p/100)/C_p\,]$ (II-6)

Upon simplification:

$S_{MB} = K_s W_o (1-C_o/C_p)$ (II-7)

3. K_s may be approximated in the following manner.

Basis: 1 incremental pound of steam flow to the steam chest of the first effect.

(a) Calculate the average rate of heat release per pound of steam condensed in the steam chest $\bar{H}_s$.

(b) Calculate the average heat of vaporization in the first effect $\bar{\lambda}_1$ at the average temperature and solution concentration.

(c) Calculate the average first effect overhead flow

$$\bar{O}_1 = \bar{H}_s/\bar{\lambda}_1$$

(d) Calculate the average rate of heat release $\bar{H}_1$ in the second effect steam chest due to the total condensation of $\bar{O}_1$.

(e) Calculate the average heat of vaporization in the second effect $\bar{\lambda}_2$ at average temperature and concentration.

(f) Calculate the average second effect overhead flow

$$\bar{O}_2 = \bar{H}_1/\bar{\lambda}_2$$

(g) $W_e = \bar{O}_1 + \bar{O}_2$ and $K_s = \dfrac{1}{\bar{O}_1 + \bar{O}_2}$

4. Although equation (II-7) describes the steam flow that will be required to concentrate a feed solution at its boiling point to the set point composition, there are two more sources of steam

consumption within the evaporator. When the feed solution temperature is below that of the first effect, it must be heated to the boiling point. The extra steam flow which is required to supply this latent heat to the feed is

$$S_{LH} = \frac{W_o(\bar{T}_1 - \bar{T}_o)}{\bar{\bar{H}}_s} \text{ with the feed heat capacity } = 1.0 \text{ Btu/lb.°F.} \qquad \text{(II-8)}$$

It is also evident that the average heat losses from the first effect (HL_1), second effect (HL_2), and inter-stage piping (HL_{12}) must be compensated for by an increase in the flow rate of supply steam. The steam flow required to offset heat losses is

$$S_{HL} = \frac{\overline{HL}_1 + \overline{HL}_2 + \overline{HL}_{12}}{\bar{H}_s} \qquad \text{(II-9)}$$

The total steam flow requirement for the evaporator has now become: $S = S_{MB} + S_{HL} + S_{LH}$ (II-10)

The derivation of equation (II-10) is quite straightforward. If does, however, illustrate how complications can very quickly arise--even in the development of a relatively simple steady state model. When dynamic compensation for feed concentration disturbances was added to the model, the amount of analog equipment required to program it became quite extensive by industrial standards. It is quite easy to foresee that as the model is extended to include more of the process dynamics, its complexity will rapidly approach the point where the model will be best suited for programming on a digital control system.

CHAPTER III

LITERATURE SURVEY

It is reasonable to propose that if the feedforward product concentration control system developed in this report is to be applied to any other evaporator, the control objectives should be similar for the two installations. Therefore an extensive search of the literature for articles pertaining to the operation and control of single and multiple effect evaporators was conducted with two objectives in mind. The first was to compare the control system and control objectives for the laboratory double effect evaporator to those reported in the literature for other commercial and experimental installations. Secondly, it was desirable to determine the points of similarity and/or difference between the existing double effect evaporator and the other evaporators for which design and experimental data has been reported.

A. Articles Discussing General Evaporator Operation

Many of the articles encountered treated the topic of evaporation very generally and gave extensive reviews with no experimental data (5, 11, 19). Two articles by Skelland (17, 18) deal with optimum operation of single and multiple effect evaporators but include only economic data.

Starczewski (20) develops and presents a graphical design procedure for a single tube evaporator. This procedure is based on the use of a correlation which predicts the boiling heat transfer coefficient at any particular point in the evaporator. He claims an accuracy of ±30% in the total heat transfer rate predicted by this correlation.

The author also presents a "possible correlation" for the effects of vapor blanketing on the performance of evaporators containing one or many tubes.

A number of articles present dynamic models developed to describe the operation of specific evaporator installations. Andre (2) describes the double effect evaporator which he designed and built to obtain experimental verification of his model. Anderson, Glasson and Lees (1) present a model for a single effect industrial evaporator while Zåvorka, Sutek, Aguado, and Delgado (21) developed a generalized model for a single effect of a commercial sugar evaporator. Their model, which they extended to a triple effect system after simplification, includes an approximation for the analytically unknown, non-linear relationship between the heating tube internal heat transfer coefficient and the juice concentration and liquid level in the calandria. They removed one variable from this relationship, by maintaining level control on the solution, and then linearized the experimentally determined relationship between the heat transfer coefficient and the juice concentration. All of the models discussed above were linearized or in some manner simplified so that dynamic response information could be more readily obtained from them. Andre's model was programmed for solution on the 7040 digital computer, the others were programmed for analog computer analysis.

Another recent article by Månczak (14) presents an analytical procedure to determine the dynamic properties of single and multiple effect sugar evaporators. The procedure was developed with the intent of optimizing the operation of an evaporator plant. Once again, extensive linearization of the dynamic equations was required around normal

operating points. Since the evaporator time constants in each effect were shown to be functions of the evaporator pressure, evaporator dimensions, solution flow rates and liquid level, the author recommended that time constants be recalculated following any significant deviation from the normal steady state operation point.

Some of the articles reviewed did include information concerning the heat transfer coefficient and time constant(s) which might be considered as typical for various types of evaporators. Carter and Kraybill (4) present heat transfer coefficients for a wide variety of evaporators operating at temperature differences of 20° and 30°F. In addition to steam and tube side heat transfer coefficients, Anderson et al (1) present the time constants for each component of the evaporator. The principal time constants for the system were 0.718 and 0.169 hours for mass and heat transfer respectively, although the volume/throughput ratio was 1.64 hours. As the authors point out, the volume/throughput ratio gives no more than an idea of the order of magnitude of the characteristic time constant. The above data are presented in Table 1 along with the corresponding information from the double effect evaporator as reported by Andre. Also included in the table are the heat transfer coefficients which have been determined from the correlations presented by Perry (15) for short-tube-vertical and forced circulation evaporators. For the latter coefficient, a fluid recirculation rate of 200 pounds per minute in the second effect (as reported by Andre) was used to calculate the fluid velocity in the heating tubes. The correlation used to calculate the heat transfer coefficient was:

$$U = \frac{490\ D^{.57}\ v^{3.6/L}}{\mu^{.25}\ \Delta T^{.1}}\ \text{Btu/hr.ft.}^{2}\,^{\circ}\text{F} \qquad \text{(III-1)}$$

Where: D = tube diameter (1.049 inches)

ΔT = steam temperature - fluid inlet temperature (30°F)

μ = viscosity (0.996 lb./ft.hr.)

v = fluid inlet velocity (4.35 ft./sec.)

L = heating tube length (5 ft.)

Quantities in brackets are the numbers used in Equation (III-1).

The shortage of information required to complete Table 1 makes it impossible to compare the various reports in a quantitative manner. It does appear that the unit operations laboratory evaporator has much better heat transfer in the first effect calandria than Anderson et al and Carter and Kraybill reported for industrial natural circulation evaporators, and slightly poorer heat transfer rates than those presented by Perry. The average heat transfer rate as calculated by Andre for the forced circulation evaporator (FCE) is considerably below the rates considered typical by Carter and Kraybill and Perry. Factors which may have caused the actual heat transfer coefficient to be lower than that predicted using equation (III-1) are possible air blanketing of the heat transfer surface and inadequate circulation through the second effect.

The ten minute volume/throughout ratio for each effect is considerably below the 1.64 hour ratio reported by Anderson et al. Reports on black liquor evaporators (reverse feed) used in the paper industry define a lag time as the time which passes before a feed flow change is detected by the concentration controller on the first effect. Kulwiec (13) reports a twenty minute lag across an evaporating unit

TABLE 1

A COMPARISON OF FORCED AND NATURAL CIRCULATION EVAPORATORS

Author	Evaporator Type	Time Constant	Heat Transfer Coefficients Btu/hr.ft.^{2}°F Steam	Tube	Overall	Temperatures Difference°F	Evaporating Temperatures°F
Andre (2)	(a) Calandria	10 min.[b]	1520	505	379	23	180
	(b) FCE	10 min.[b]			358	42	150
Anderson et al (1)	Calandria	0.718 hr.	246	81	60.4[a]	35	200
Perry	(a) Calandria				375	20	180
					450	30	180
					490	50	180
	(b) FCE				1155	30	150
Carter & Kraybill (4)	(a) Calandria				40	20	100
					220	20	140
	(b) FCE				450	20	100
					530	20	140

[a] Calculated

[b] Volume/throughput ratio

The evaporators referred to as calandria type are short-tube-vertical evaporators with natural circulation: FCE refers to forced circulation vertical evaporators.

containing four forced circulation long-tube-vertical evaporators (LTVE) and a soap separation tank, and a seven minute lag across the first two effects. Referring to the composition response curves presented by Andre [(2), p. 97] for the unit operations evaporator, it can be seen that composition changes are almost immediately detectable in both effects after changes in feed flow rates.

On the basis of this very qualitative comparison of the unit operations double effect evaporator to some of the larger industrial evaporators which have been described in the literature, a few observations and comments may be summarized here regarding the double effect evaporator.

1. The first effect performance is comparable to that normally expected from a natural circulation evaporator.
2. The performance of the second effect is below normal: the heat transfer coefficient in a forced circulation evaporator should be greater than that of a natural circulation evaporator to justify the expense of operating a recirculating pump.
3. The evaporator system appears to be very sensitive and responds to disturbances in its input variables much faster than do the larger industrial evaporators. It should therefore be quite easy to control the laboratory evaporator as if it were a single unit. To illustrate this supposition, it can be pointed out that the successful control scheme worked out by Andre controlled the product concentration at the last (second) effect by regulating the steam flow to the first effect. For commercial evaporating systems this type of control is usually avoided

by (a) controlling product concentration via regulation of the product flow from the last effect if a forward feed evaporating system is used or (b) using a reverse feed system so that the product concentration can be controlled via manipulating the flow of supply steam to the same effect from which the product leaves. Two control schemes tested by Andre involved product concentration control via manipulation of the product discharge rate [(2), pp. 61-64], but were unstable for the controller settings investigated.

There are, of course, other reasons for using a reverse feed system and other possible means of controlling it. It might therefore be expected that when the dynamics of two evaporating processes differ significantly, the best control schemes for these processes may bear little resemblance to each other.

B. Evaporator Control and Instrumentation

The literature contains a variety of articles on the subject of evaporator control, and on the proper choice of instrumentation and control action to best achieve the objectives laid down for the operation of an evaporator. If the objectives of the control system are not kept clearly in mind it is easy to become confused by the apparent contradictions in some of these articles or by the contradictory results reported for control systems which appear to be similar. It should be remembered that the laboratory evaporator control system was designed to hold product composition on specification for a maximum amount of operating time, without using feed flow as a manipulated variable.

Since the control philosophy for single effect evaporators can often be extended to multiple effect systems with only the addition of level controllers on each effect to maintain local liquid levels, the control systems commonly encountered for single effect evaporators will be discussed first.

1. Single Effect Evaporators

The two primary controlled variables in an evaporation process are usually evaporation rate and product composition. In addition to these variables, it is usually necessary to control liquid level so that a continuous material balance is satisfied, and evaporator vapor pressure (or temperature) so that maximum steam economy is achieved. The controllable variables are heat supply, evaporator body pressure, cooling water supply rate, and feed and product flow rates. Clearly, these variables are not all independent of one another and cannot all be controlled simultaneously. For example, if the feed rate is fixed and any two of the first three controllable variables listed above are fixed, product flow rate and the remaining variable of the three must be determined.

All of the stable control systems discussed in the literature manipulated the product discharge rate to achieve either concentration or liquid level control. If it is assumed that product rate must be controlled for one of these purposes, feed or steam flow rate can be regulated to achieve the other objective. Thus four different control configurations are possible for a single effect evaporator (Table 2).

The fourth control configuration was not reported in the literature for a single effect, but was not capable of providing stable product concentration control for the laboratory double effect evaporator [(2), p. 61].

Table 2

CONTROL OF A SINGLE EFFECT EVAPORATOR

Configuration	Steam Rate Set By:	Feed Rate Set By:	Product Rate Set By:	See Figures
I	-	Concentration Controller	Level Controller	2.1, 2.5
II	Concentration Controller	-	Level Controller	2.2, 3.1, 3.2
III	-	Level Controller	Concentration Controller	2.3, 2.4, 3.3
IV	Level Controller	-	Concentration Controller	-

Using an analog computer, Anderson, Classon and Lees (1) simulated the response of an calandria-type evaporator using six different control systems. The three stable control systems which included product concentration control are illustrated in Figure 2. Operating pressure was determined by cooling water flow rate and was assumed constant. This left feed, steam or product flow rates free for manipulation to control product composition and liquid level. They noted that:

(a) With the evaporator discharge under level control, the composition controller may act directly to control feed (Figure 2.1) or steam rate (Figure 2.2). Better composition control resulted when feed flow was manipulated by the composition controller.

(b) Attempts to control product composition via steam regulation and liquid level by the regulation of feed flow led to stabilization difficulties.

(c) Control of product composition via regulation of product discharge, and level control via feed rate regulation

(Figure 2.3), results in indirect composition control since the composition controller must act through the level controller. The response of this system was rapid, but the amplitude of oscillations in product concentration induced by a step change in steam flow was quite large due to interaction between the two control loops.

The only scheme not using feed as a manipulated variable (Figure 2.2) was found to give somewhat inferior composition control; this scheme is similar to that used by Andre on the double effect evaporator.

In a general discussion of evaporation, Standiford (19) states that with evaporation rate constant, (or determined by steam flow rate) concentration responses to load and set point changes are more rapid when composition control manipulates the discharge valve and level control manipulates evaporator feed (Figure 2.4), rather than when the above manipulated variables are reversed (Figure 2.5). Since Figures 2.1 and 2.5 and 2.3 and 2.4 are identical, this observation should agree with that of Anderson et al. Both articles state that the response of control scheme 2.3 (or 2.4) is very rapid following system disturbances, but Anderson et al qualify their observation by showing that it is also more oscillatory than the responses obtained using the control schemes of Figures 2.1 and 2.2.

Drinker and Barber (6) originally discussed the topics reviewed by Standiford. They state that although manual startup is required for an evaporator using control scheme 2.5 as opposed to a possible automatic startup using scheme 2.4, the controlling of feed rate by composition control and discharge rate by level control allows a greater amount of product to be produced closer to specifications during

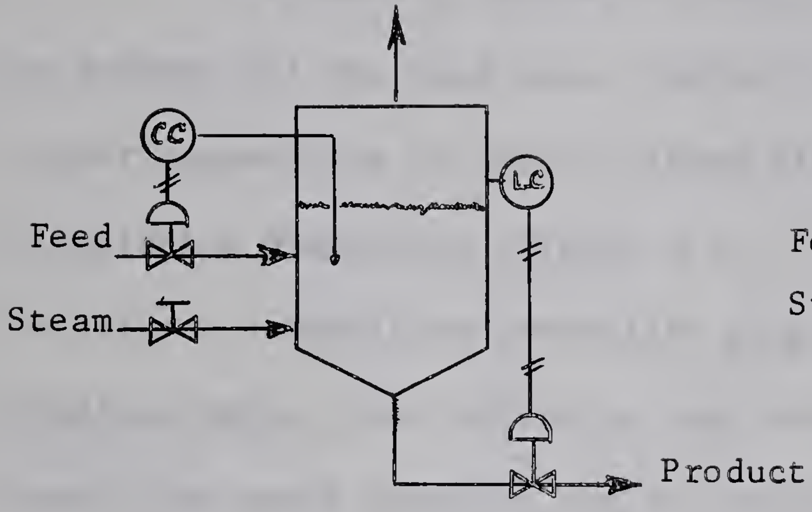


Figure 2.1
Anderson, Glasson, & Lees (17).

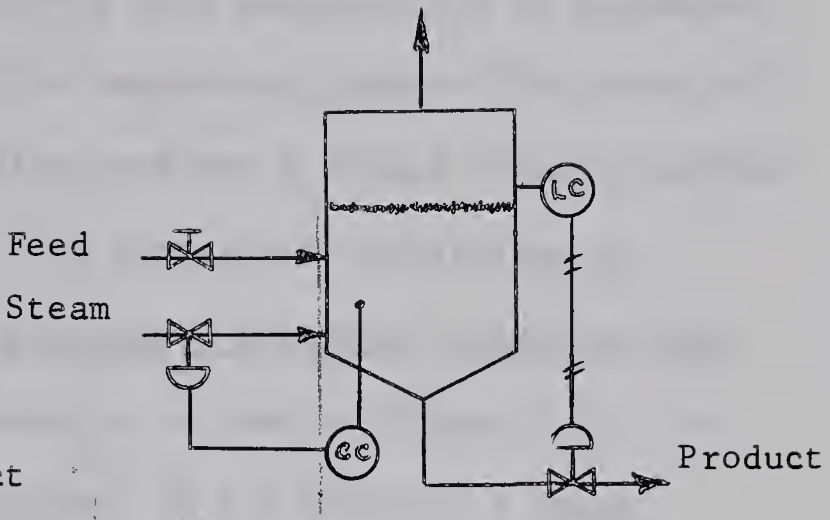


Figure 2.2

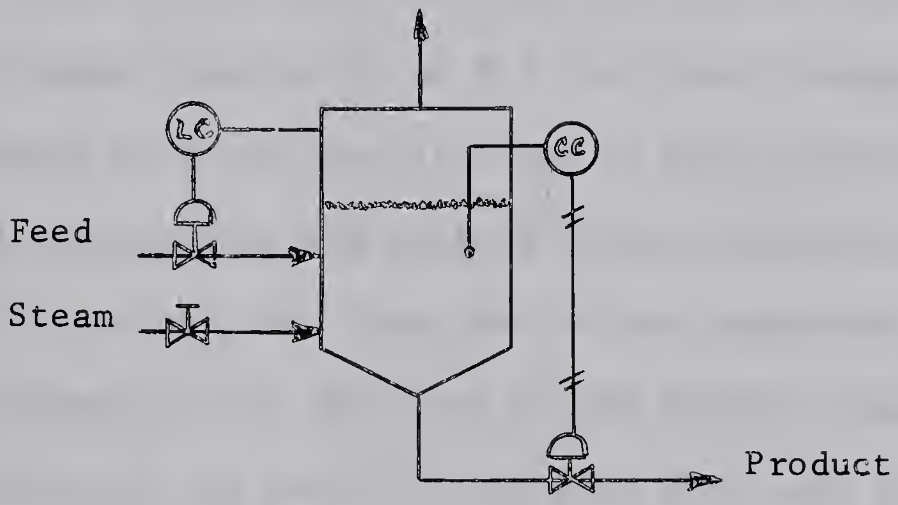


Figure 2.3

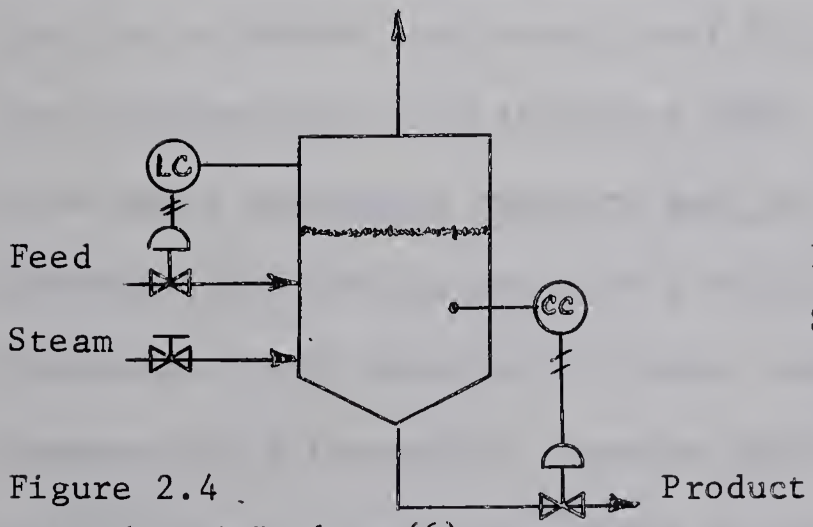


Figure 2.4
Drinker & Barber (6)
Standiford (19)

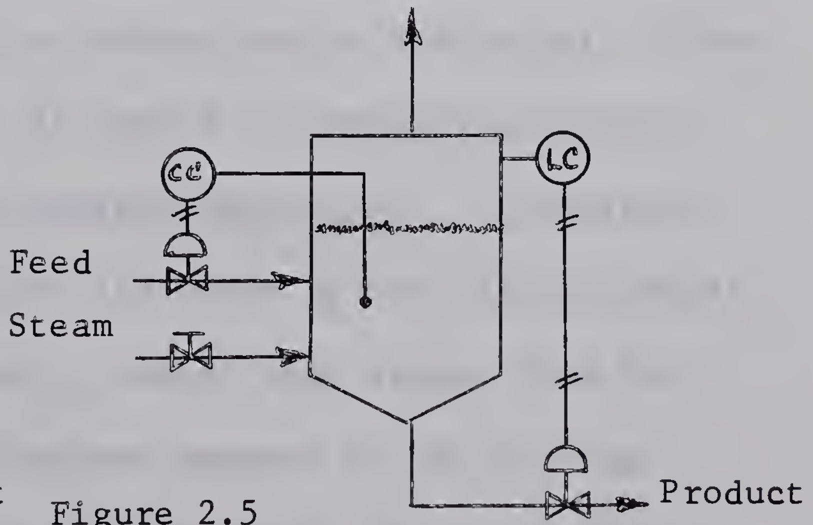


Figure 2.5

Figure 2 CONTROL SCHEMES FOR SINGLE EFFECT EVAPORATORS
A Summary of Results Reported in the Literature.

initial operations. This observation is in good agreement with that made by Anderson et al for the same control system (Figure 2.1).

A control scheme not requiring feed manipulation is presented by Johnson (9) who used level control to manipulate product flow rate and liquor temperature to control steam flow rate for a single effect, natural circulation evaporator (Figure 3.1). The temperature controller is actually a composition controller if evaporator pressure is held at one absolute value; this system is then similar to that of Figure 2.2. To supply the exact pressure control required, an air bleed to a steam ejector was controlled. Johnson noted that there is a very strong interaction between vacuum and liquid level in this system. A 0.1 inch of mercury pressure change resulted in an 8.5 inch level change in the evaporator. Although no other data is given on the behavior of this evaporator, it is evident that the control system performed satisfactorily and did not have to control feed flow rate to the evaporator. A similar control system (Figure 3.2) is described by the Foxboro Company (24). It has a flow controller on the feed, but the feed flow rate is not dependent upon any of the evaporator variables. In this case steam pressure is part of a cascade loop being reset by the concentration controller. Since the boiling-point-rise technique (BPR) is used for composition control, once again evaporator pressure must be closely controlled. At constant pressure, the boiling point of a solution increases as its solute content increases. This increase is linear over a fairly wide range; thus the temperature differential, between temperature sensors in the boiling solution and in vapor condensing at the same pressure, is an indication of the solution concentration. For this system pressure is controlled by

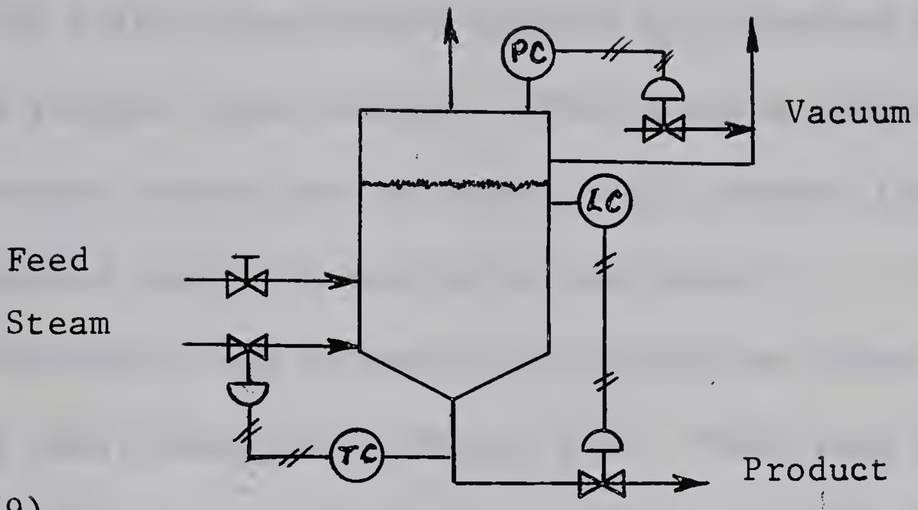


Figure 3.1
Johnson (9).

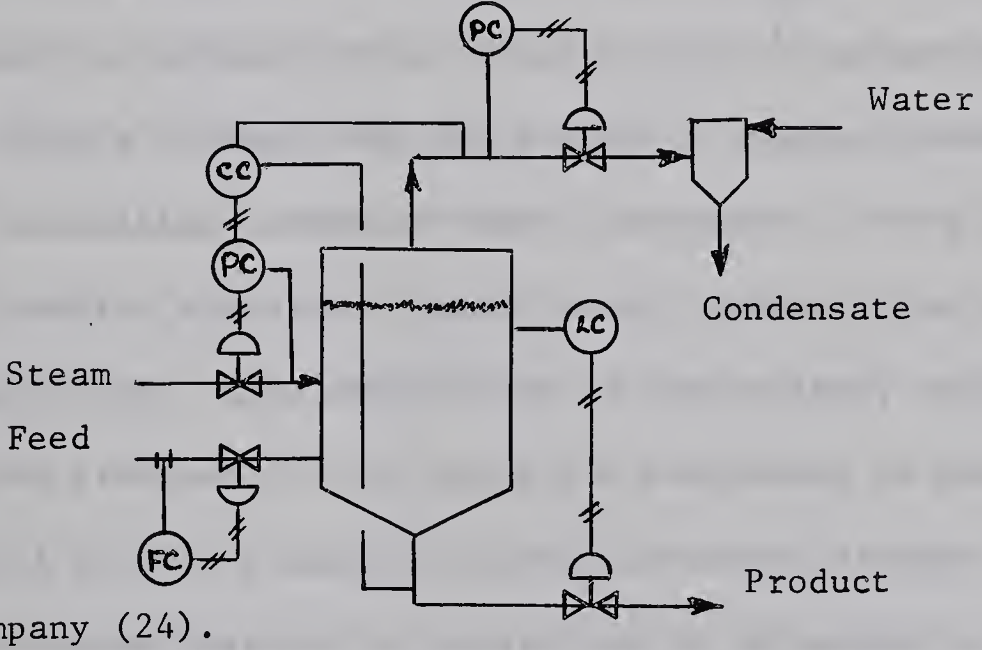


Figure 3.2
Foxboro Company (24).

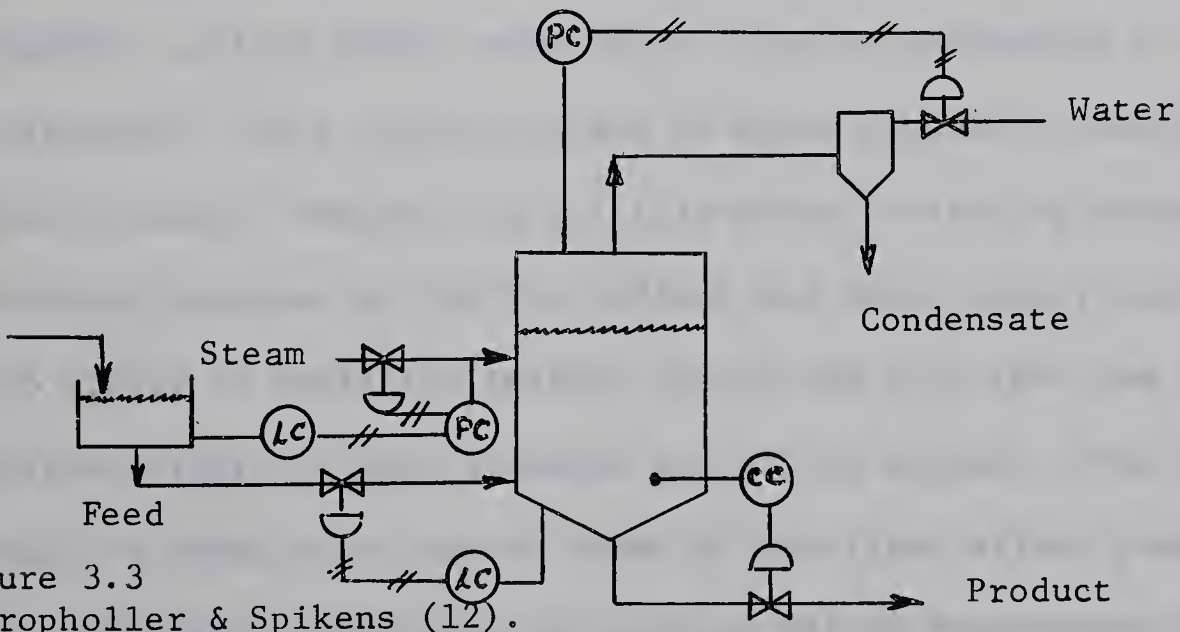


Figure 3.3
Kropholler & Spikens (12).

Figure 3 CONTROL SCHEMES FOR SINGLE EFFECT EVAPORATORS (Continued)

manipulation of a butterfly valve in the overhead vapor line. This technique results in good pressure control, but is not usually considered for use in multiple effect evaporators because the pressure drop across the control valve reduces steam economy. The bulletin recommends that proportional-reset controllers be used on all control loops, but no information is included regarding controller settings.

Kropholler and Spikens (12) include an inventory tank in the feed line for their evaporator (Figure 3.3). This tank helps stabilize the operation of the evaporator by dampening out smaller disturbances in feed flow rate and concentration. They propose to accept feed at variable rates from a storage tank and produce a constant composition product, while minimizing condenser water consumption. Feed flow is manipulated to control evaporator liquid level: product flow is throttled to control composition. This combination of controllers, unlike the two systems described previously, will allow the evaporator to start-up under automatic control without producing off-specification product. Since evaporator vapor-space pressure is controlled by adjustment of the flow of condenser cooling water, evaporation rate is determined by controlling steam pressure. This control scheme is often applied to multiple effect evaporators also. However, in multiple effect units the dynamic relationship between pressure in the last effect and steam supply pressure is usually slower to establish itself, due to the fact that the resistances each effect offers to heat transfer are all in series. Thus flow control of steam, or steam flow control reset by the first effect pressure controller is more often encountered on multiple effect evaporators when evaporation rate is to be controlled. For the single effect evaporator

Kropholler and Spikens choose to vary evaporation rate with feed rate to minimize feed storage. This was done by transmitting a level control signal from the feed tank to reset the steam pressure controller set point. Level control on the feed tank is therefore quite slow since an adjustment to the evaporation rate must first be followed by an evaporator level change. This level change then adjusts the feed flow rate exercising level control on the feed tank. Although minimum water consumption is obtained when the condensate temperature is controlled by manipulating cooling water flow, the authors suggest that if pressure fluctuations are to be avoided during evaporation rate changes, the flow of condenser cooling water should be set by the pressure controller. Since the level controller acts on a first order system, proportional control is considered adequate. Similarly, proportional control is suitable for steam pressure control since this is the inner loop portion of a cascade system. Proportional-reset control action is recommended for pressure control, feed tank level control and product composition control. (This last loop may also require derivative action if a sampling-type composition controller is used.) The authors recommend the Ziegler - Nichols undamped oscillation technique (22) for the calculation of controller settings for all control modes.

2. Multiple Effect Evaporators

The reverse feed evaporation system, (Figure 4.1) as described by Drinker and Barber (6) appears to be the most common method of operating industrial evaporators. The single effect control scheme of Figure 2.5 cannot be extended to reverse feed, multiple effect evaporators as it would be impractical to control product concentration via regulation

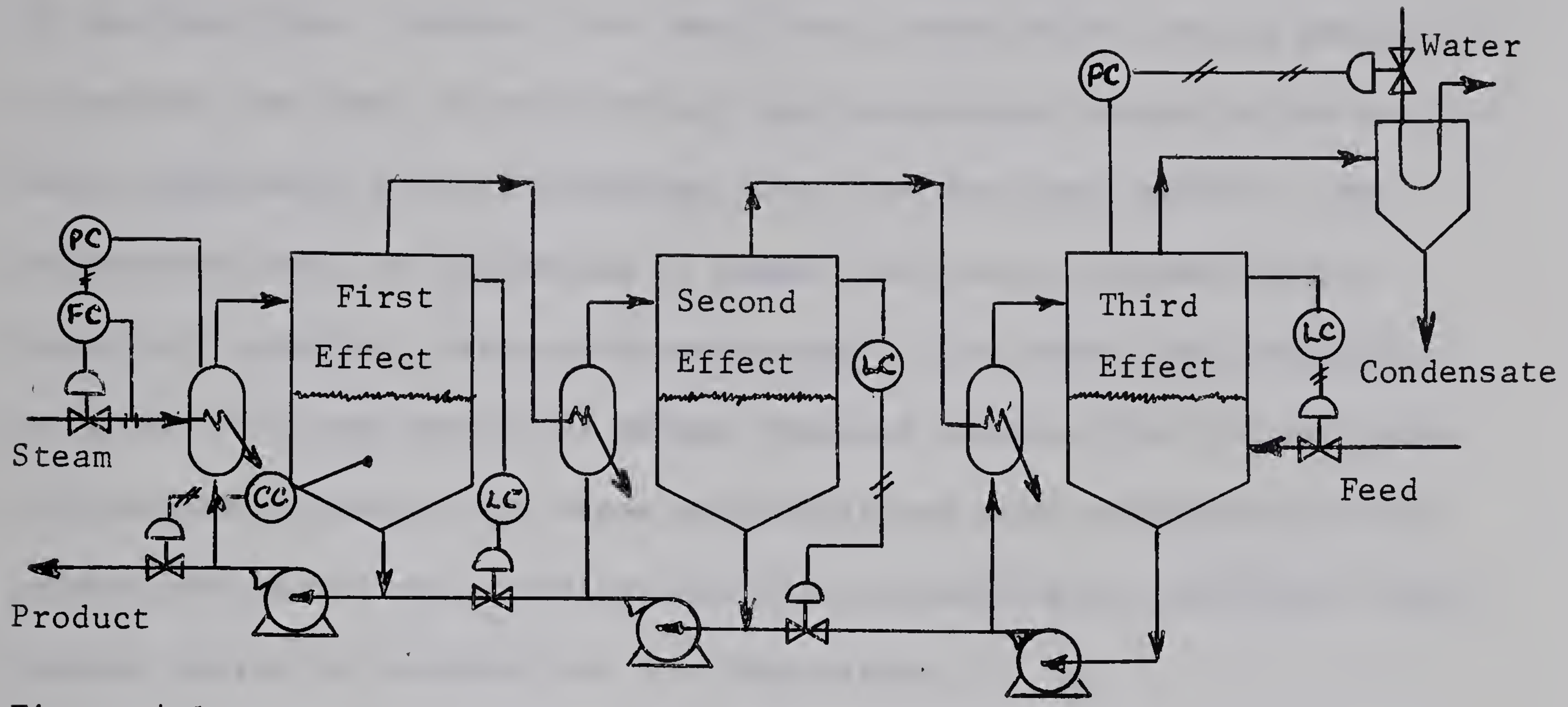


Figure 4.1
Drinker & Barber (6).

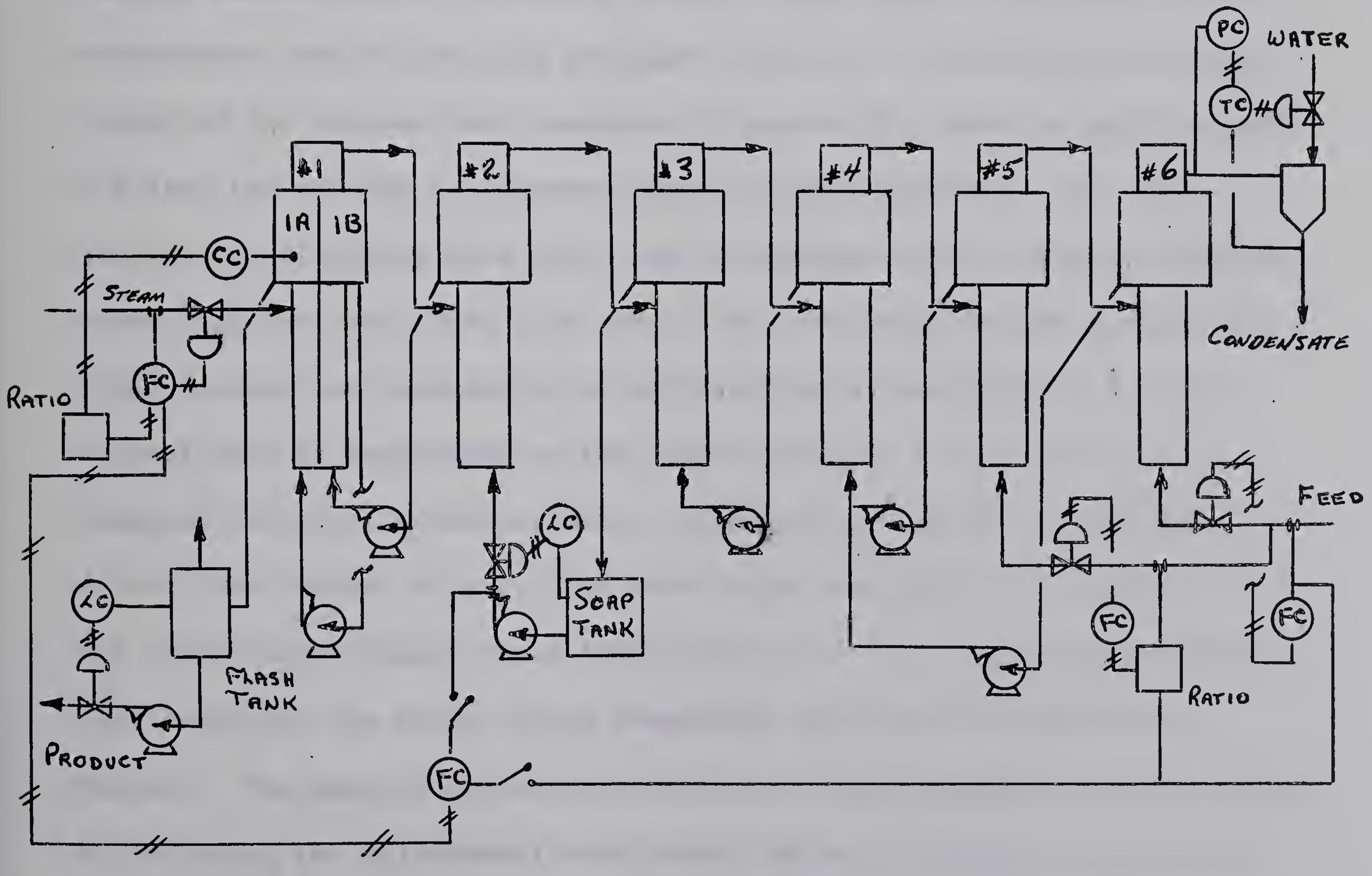


Figure 4.2
Howard (8).

Figure 4 CONTROL SCHEMES FOR MULTIPLE EFFECT EVAPORATORS

of the feed flow. Instead, the feed flow to each effect may be manipulated to control the level in that effect, and the product concentration controlled by regulating product discharge rate from the first effect. The evaporation rate, as determined by steam flow rate, is usually held relatively constant, with adjustments made to the steam flow controller set point only when the first effect pressure departs from its set point. This cascade signal to the steam controller may also originate from the product concentration controller (8, 13), in which case a different level control system is required for the first effect.

Drinker and Barber (6), Howard (8) and Kulwiec (13) describe control schemes for the multiple effect, black liquor, long-tube-vertical evaporators used by the pulp and paper industry. Six effects are usually installed for reverse feed operation (Figure 4.2). Feed is split between the last two effects to increase the evaporator capacity. The flow controllers effecting this split are independent of all process variables other than the total feed flow rate. The evaporator design is such that level control instrumentation is not required at each effect. Product removal rate is determined by the liquid level in a flash tank, which receives the first effect product. Composition control is maintained at the first effect by a boiling-point-rise controller which resets the set point on the supply steam flow controller (or on the steam pressure controller for the black liquor evaporator described by Drinker and Barber). The control action of this loop is quite gradual as can be seen by following the adjustments which result after a change in liquor concentration reaches the first effect. This change causes an adjustment to be made to steam supply pressure and flow rate. The effect of this change

in heat input is not evidenced immediately by the liquor concentration, since heat transfer resistances and system heat capacities delay the evaporation rate adjustment called for by the composition controller. As the evaporation rate in the first effect adjusts to bring the BPR to the set point, each downstream effect undergoes a pressure change as it seeks a new equilibrium for heat transfer. This will change the liquor concentration in each effect with the result that a new disturbance in solution concentration will reach the first effect after some time. It is evident that if stable operation of the evaporator is to be achieved, the system correction for the initial upset must not be too rapid or too severe. The large capacity of this system, combined with the fact that feed fluctuations may not be felt by the composition controller before twenty minutes elapse makes control of the unit very difficult. Overcompensation was a serious problem when steam flow was manually regulated in an attempt to provide product concentration control (13). Howard (8) attempted to hasten steam flow compensation for feed flow variations by using a steam-to-feed ratio controller (Figure 4.2). This "anticipator control" system still had an 8 to 10 minute lag before a feed upset (and the corresponding steam flow adjustment) was actually detected by the concentration controller. Therefore he used an additional ratio controller to control steam flow at a preset ratio to the liquor flow to the second effect: the feed flow ratio controller was made secondary to be used only during changes of load capacity. The BPR controller continuously determined the set point for the steam-to-liquor flow ratio controller. To achieve maximum water evaporation with a minimum consumption of steam, and maintain constant product composition, close absolute pressure control

on the last effect was essential. To achieve this pressure control with a minimum consumption of cooling water, Howard measured vapor pressure in the last effect and condensate tail pipe temperature. These signals were used to continuously compute the flow of cooling water to the barometric condenser. The success of Howard's anticipatory control for steam flow (1962) led Kulwiec to install a similar control system on four black liquor evaporators which had been controlled manually (1965). He reported that close composition control was achieved and that the lag time before feed rate disturbances were detected by the BPR controller was cut from 20 to 7 minutes.

The control system developed by Andre for the laboratory evaporator is in many respects analogous to that described above for the black liquor evaporator. Although all of his findings were based on a forward feed evaporator system as opposed to the reverse feed system used for the black liquor evaporator, Andre's conclusions with respect to evaporator performance indicate that his control scheme relies upon inherent evaporator characteristics similar to those of the black liquor evaporator. To illustrate, his conclusions were:

1. Steam flow changes to the first effect calandria have a greater effect on the second effect level and composition than on the first effect level or composition. Of all these variables, second effect product composition is most sensitive to steam disturbances.
2. Disturbances originating at the first effect are transmitted to the second effect through the vapor and solution flows leaving the first effect. The disturbances transmitted through the vapor phase are detectable more rapidly and to a greater extent, by product concentration changes, than are liquid phase disturbances.

3. The response of the second effect concentration to steam flow changes is almost as fast, and is more noticeable than the response of concentration in the first effect.

His successful control scheme therefore attempted to hold the most important variable (product composition) on the set point for a maximum amount of time by manipulation of steam flow rate--the variable which has the greatest effect on product composition. Steam flow changes are evidenced almost immediately by corresponding pressure changes in the first effect and heat transfer rate adjustments in both effects. Since second effect composition responds nearly as quickly as concentration in the first effect, and changes to a greater degree than the first, it was logical that steam flow be the variable manipulated to control product concentration. This concentration control system should work even better for a reverse feed system, since the manipulated variable (steam flow) and controlled variable (product composition) would then both be at the first effect as in Figure 4.2.

These observations point out the possibility of regulating inter-stage vapor pressure, via the addition (or venting) of steam at the steam chest of the product effect, to provide an alternative for product concentration control. This alternative would separate the product effect from the remainder of the multiple effect unit and must result in a decreased steam economy. The inter-stage pressure might also be controlled via resetting the steam flow rate to the first effect, but feed flow rate would then, of necessity, become a manipulated variable as in Figure 4.1.

The most serious problems evaporator control systems must face are the large system time constants and transportation lags which distort

and delay the response of product concentration to feed flow and concentration upsets. No feedback control system can compensate for these upsets until they appear as errors in the measured product concentration. Fortunately, the time constants and transportation lags were quite small for the unit operations double effect evaporator, and did not seriously handicap the feedback control system developed by Andre.

Howard and Kulwiec attempted to alleviate these problems by using anticipatory control which effectively decreased the measurement lag. The feedforward control system designed for this project determines the steam flow requirements using a model which measures only system input variables and set points. Therefore, it is hoped that its compensating adjustments to the steam flow rate will succeed in offsetting at least a major portion of the effects of feed disturbances before they are detectable in the product concentration. Negative feedback control will be retained as an option within the feedforward control system to ensure that product composition will not drift from the set point under changing load conditions downstream of the first effect calandria.

CHAPTER IV

EVAPORATOR PROCESS EQUIPMENT

A. Evaporator Construction

The major components of the evaporating system designed by Andre remained unchanged during this study. They are:

1. first effect; natural circulation, calandria-type evaporator,
2. second effect; forced circulation evaporator,
3. vapor-liquid cyclone separator,
4. overhead total condenser,
5. condensate run-down tank.

Andre's drawings, photographs and discussion of these components and their supporting framework [(2), pp. 48-55] provide an accurate description of the basic evaporator system, which will not be repeated in this chapter.

A number of modifications and additions were made to the process to improve evaporator performance, permit easier calibration of flow transmitters, and remove some of the noise which was reported by Andre as inherent in many of the process variables [(2), p. 80].

1. Evaporator Performance

The operating difficulties which most affected the efficiency of the evaporating unit were located in the second effect recirculation and in the condenser condensate removal systems.

(a) Second Effect Recirculation System

The low average heat transfer coefficient in the forced circulation evaporator (358 Btu/hr.ft.^{2}°F as compared to the average of 379 Btu/hr.ft.^{2}°F for the first effect) reported by Andre did not appear normal or acceptable in view of the extra

operating expenses encountered in pumping the solution through the second effect. It appeared that this low heat transfer coefficient might have been the result of a low recirculation velocity through the heating tubes. The model 1L4 Moyno pump has a rated maximum flow of 200 pounds per minute at zero discharge pressure (26) and maximum speed. The pump discharge line was a tortuous section of 0.50 inch pipe containing a control valve, a refractometer and an in-line flow transmitter. The equivalent length of this line was at least 40 feet. With a flow rate of 200 pounds per minute, the friction pressure drop alone would be well above the 40 psi differential within which the pump was to operate. Under these conditions it is estimated that the maximum circulation rate could not exceed 130 lbs./min. (or 3.56 feed per second in the heating tubes). Perry (15) states that normal forced circulation velocities should be in the range from 4 to approximately 10 ft./sec. if good heat transfer is to be achieved. In an effort to raise the circulation velocity, the discharge line was replaced by a 1.50 inch copper pipeline which included only an orifice run and the product refractometer. The pump differential pressure under these conditions (8 psi) was inadequate to drive product through the product line to tankage. The circulating fluid also flashed below the orifice run causing chatter in the line. A 1.50 inch globe valve was then installed in the recirculation line to hold the pump discharge pressure sufficiently high (approximately 20 psig) to allow controlled product removal, and to prevent flashing in the line. Under these conditions the Moyno pump cavitated continuously and caused noise in both the recirculation and first effect bottoms flow signals.

The Moyno pump was therefore replaced with a Goulds model 3197-STD centrifugal pump. This pump will deliver 50 USGPM at 10 psi differential when operated at 1750 RPM. The corresponding heating tube velocity at this circulation rate would be 9.05 ft./sec. Since the suction pressure was normally only 4.5 psia, the discharge pressure was not high enough to drive product from the recirculating line to the product tank. A small Jabsco model JC 1/2 pump was therefore installed below the Goulds pump to drive product to tankage. Although this pump performed favorably for the duration of the experimental runs, it did develop a small solution leak through the stuffing box after it had been in use for some time. It was driven at very low speed by a variable speed motor in an attempt to prolong the life of the seal. Figure 5 is a schematic diagram of the new recirculation system and its associated instrumentation. This system performed very satisfactorily, and resulted in an increase in the heat transfer coefficient at the second effect to an average value of 461 Btu/hr.ft.^{2}°F.

(b) Condenser Condensate Removal

During early evaporator test runs, condenser flooding was noted when the condenser pressure approached the vacuum pressure in the run-down tank. This was attributed to an insufficient differential pressure at the level control valve in the vacuum line. The condensate level control loop was not essential to the operation of the condenser and served only to maintain an inventory of condensate above the thermocouples and the flow transmitter in the condensate line. It was therefore removed and replaced by a spillover level control system which held just enough water in the condenser to keep the condensate thermocouples

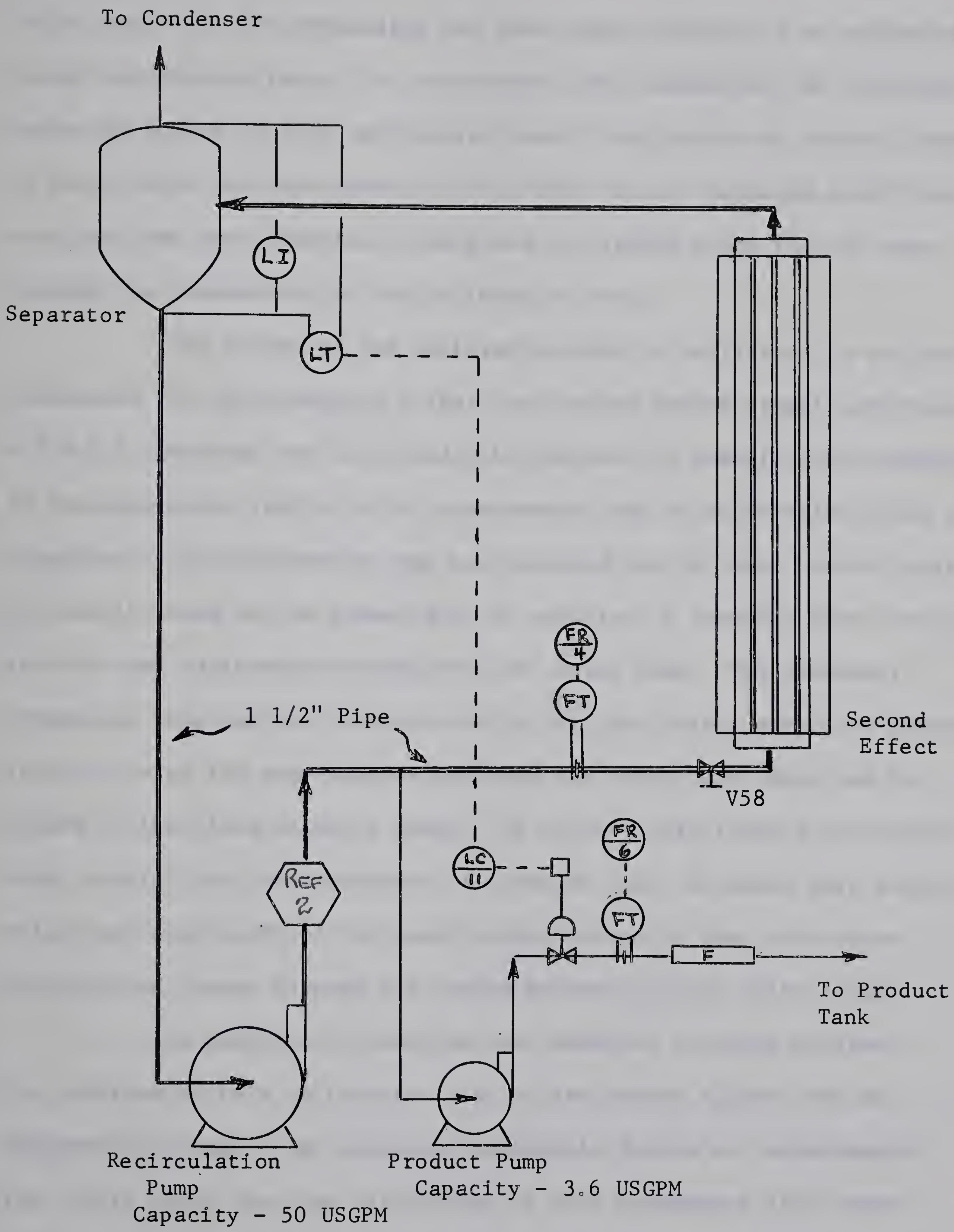


Figure 5 SECOND EFFECT RECIRCULATION SYSTEM

[illegible]

[illegible]

fully immersed. The condensate was then piped directly to an evacuated 40 gallon calibration tank. The condensate flow transmitter was relocated below the bottom of this calibration tank. Thus there was always a head of water above the flow transmitter. Under steady operating conditions, this head was just adequate to maintain the steady state flow of water through the transmitter to the calibration tank.

The volume of the calibration tank is sufficient to collect condensate for approximately a four hour period during normal operation. A 3/4 H.P. Dynapump was installed with the tank to transfer the condensate to the main water tank after an experimental run or calibration check was completed. The calibration tank was equipped with a level control system so that it could not be pumped dry; in addition, a constant level could be held when calibration checks were not being made. The condenser condensate pump was not usually left on for this latter operation however, since it is of the same type as the feed and auxilliary pumps and is cooled by the fluid which it pumps. If it were left running to provide level control for the condensate calibration tank, it would very rapidly raise the temperature of the small holdup volume to the point where vaporization losses through the vacuum system would be quite large.

In addition to removing the condenser flooding problem, the addition of this calibration tank to the system allowed the two condensate streams to be collected separately during any experimental run. This meant that the calibration of both condensate flow transmitters could be checked on-line during any steady state period of operation. A second glass lined calibration tank and all the associated instrumentation were installed in a portable module with the condenser

condensate tank (see Figure 6). The second tank was used to collect the flow of steam condensate from the first effect steam chest. This tank was normally drained to the sewer following completion of calibration. Both segments of the condensate collection module worked very well and were used extensively for checking the calibration of the steam and condenser condensate flow transmitters.

(c) Calibration Measurements

A further improvement to the evaporator was the installation of block and drain valves in the first effect bottoms line below the flow transmitter. (See Figure A-1, valves V34, V35.) This installation permitted calibration checks to be made on the first effect bottoms flow transmitter at operating conditions.

(d) Process Piping

The auxilliary pump had originally been connected to each of the feed, product, and water tanks with 0.5 inch plastic pipe. This pipe did not prove satisfactory. Air leaks around pipe unions and at the pump intake frequently caused cavitation in the auxilliary pump. Initially a portion of this pipe was replaced with 0.5 inch stainless steel tubing; somewhat later the remainder was replaced with copper tubing. Both types of tubing were totally acceptable and allowed smooth operation of the auxilliary pump. Although Andre mentioned that the use of copper tubing had not been totally satisfactory on some equipment in the unit operations laboratory [(2), p. 50], it appeared to be quite acceptable for the sugar-water and water duty on the evaporator. Unless further evidence indicating the inferior properties of copper tubing is dectected when the evaporator is dismantled, it is felt that the extra expense of using stainless steel

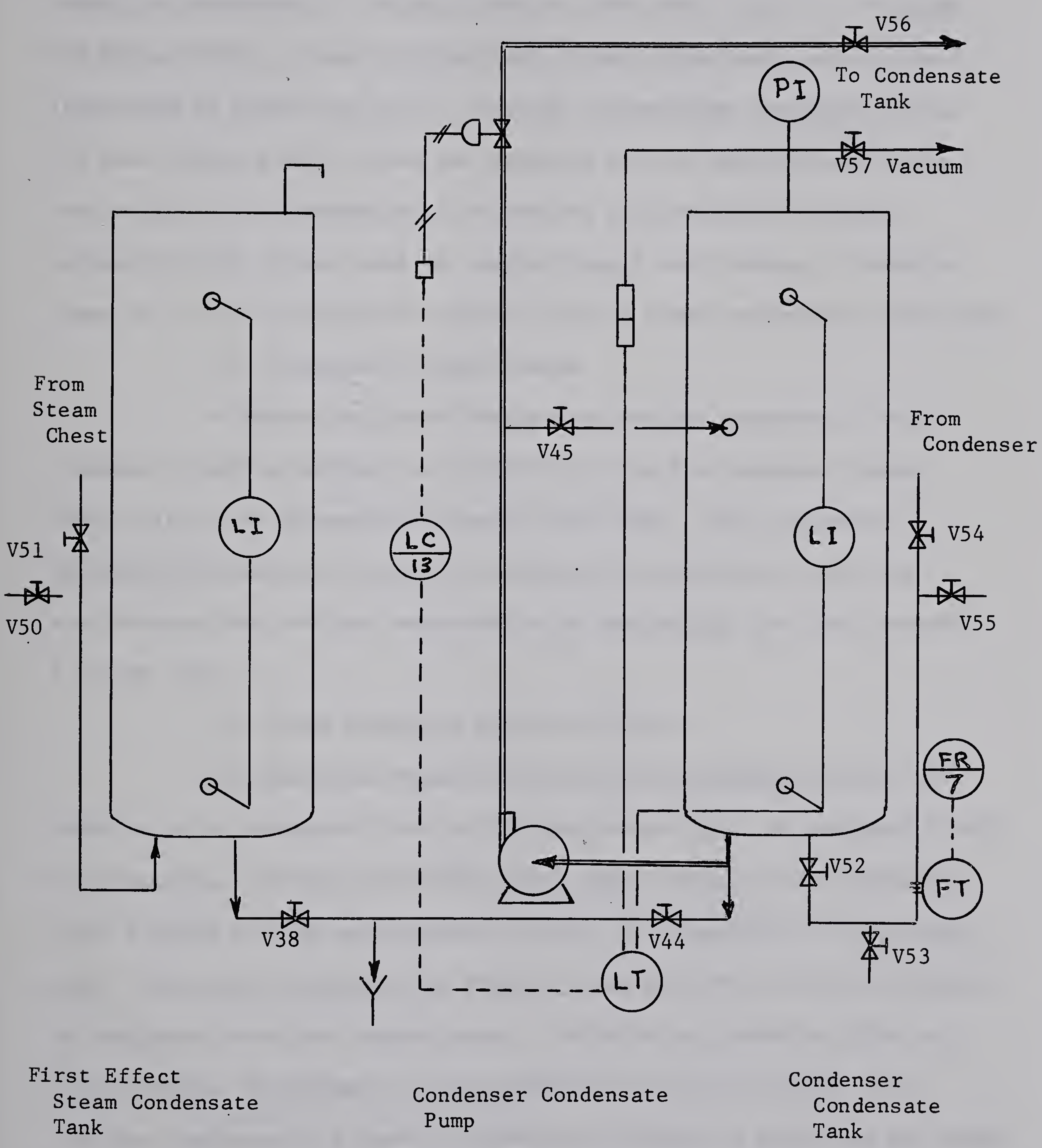


Figure 6 CONDENSATE COLLECTION SYSTEM

tubing is unnecessary. During a service check when a 3/4 H.P. Dynapump was being tested, it was observed that a pump intake any smaller than 1 inch tends to starve the pump. Although the Dynapump throughput using 0.5 inch tubing suction lines was adequate for all applications during this study, it is recommended that they be re-installed in the new laboratory with intake lines no smaller than 1 inch tubing. Discharge lines of 0.5 inch tubing are adequate for all normal evaporator flow rates.

(e) Evaporator Vacuum System

A further equipment change involved the transfer of the condenser pressure control valve (PRCV-22) from the condenser vacuum supply line to an atmospheric bleed to this line. This relocation removed the necessity of manually setting the atmospheric bleed and also avoided the problems associated with controlling the flow through a vacuum line.

(f) Steam Supply to the First Effect

The remaining change to the process equipment involved the addition of a condensate trap in the inlet steam line. It appeared likely that the steam reaching the first effect might contain some condensate, after flowing through approximately 50 feet of insulated 0.5 inch copper tube. The trap, as depicted in Figure 7, was made from a 24 inch section of insulated three inch copper pipe. The inlet was below the trap outlet to prevent entrainment of the condensate in the dry steam. The trap was continuously purged to prevent the buildup of steam and the steam pressure disturbances which often result during intermittent purging. The condensate trap was installed following Runs 20L, 30L and 40L. When the condensate trap drain valve (V62) was shut and then opened after

approximately 30 minutes, a noticeable amount of condensate was purged to the sewer. The trap was therefore a valuable addition to the evaporator system and provided a smooth flow of dry steam to the first effect.

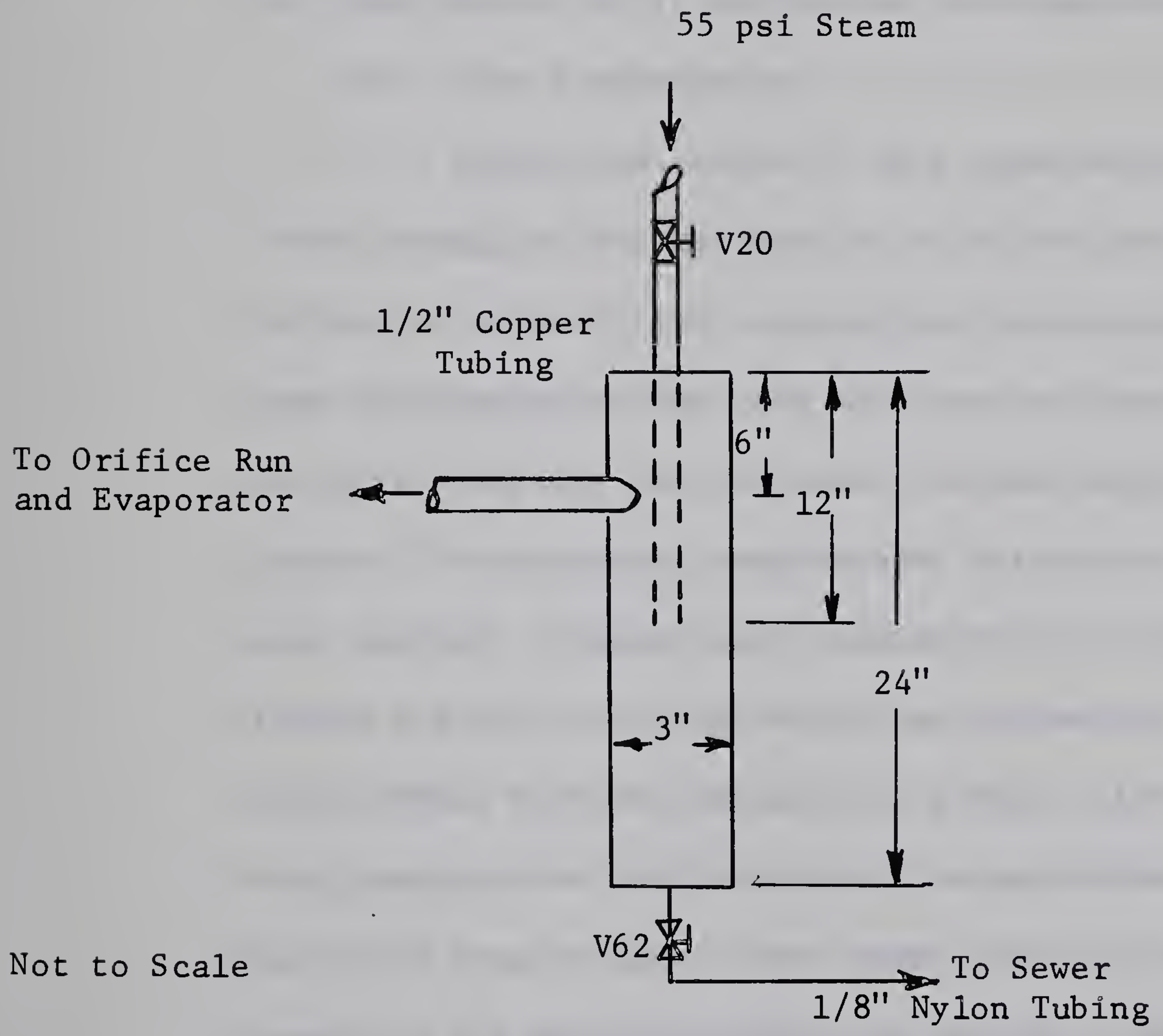


Figure 7   STEAM CONDENSATE TRAP

2. Evaporator Instrumentation

The feedforward concentration control system was developed for use with and comparison to the basic evaporator control system used by Andre. The functions of the individual control loops on the evaporator were unchanged during this study, although a number of flow transmitters

and valves were removed and/or relocated for the reasons already stated. The problems mentioned by Andre--noisy flow and level signals and drifting of refractometer and flow transmitter output signals--were also investigated in the hope of overcoming them or, at least, reducing their gravity.

(a) Calibration Drift in Process Instrumentation

(i) Flow Transmitters

During the course of this experimental work, no day-to-day change in the calibration of either the first effect bottoms or product flow transmitters was encountered. The same calibration curves were used over a three month period and their validity was confirmed intermittently by calibration checks. The operating temperatures within the dp cells were also checked. Because each transmitter is in-line and the fluids are hot, its temperature may become quite high. The first effect bottoms flow and the product flow transmitter head temperatures were determined as approximately 160°F and 140°F respectively. Both these temperatures are close enough to the maximum temperature rating of the amplifiers in the dp cell (180°F) to merit removal of the dp cells from the flow measuring device. If orifice runs, similar to that constructed for the recirculation line, are installed in these two lines the dp cells can be operated at ambient temperatures and there will be no danger of overheating their amplifiers.

(ii) Concentration Measuring Refractometers

A great deal of time and effort was spent trying to determine the cause of, and the solution for, the

refractometer drift problem. Since the temperature compensating thermistors were not rated for the operating temperatures at which they were to be used, they were removed and replaced by resistors of resistance equal to that of the thermistors at their specified mean operating temperature (7K at 200°F). This substitution removed most of the noise from the refractometer output--which appeared to be stable at room temperature. The removal of the temperature compensation option from the refractometers did not handicap the feedforward control system in any way. The product temperature was always constant since pressure in the second effect was very closely and accurately controlled. The feed temperature was to be held constant by adjusting the cooling water flow rate to a heat exchanger in the feed line. When it became evident that manual temperature control could not maintain a constant feed temperature, the feed concentration was simply determined by analysis and then the correct refractometer output was set directly within the feedforward controller on a potentiometer. Once this procedure was decided upon, the feed refractometer was theoretically required by the control system but practically, it was not used.

Although the refractometer drift was cut down considerably from the rate which was observed when the first trial runs were made on the evaporator system, it was not eliminated. Thus an average drift rate of approximately

-0.10% sugar per hour was still inherent in the product refractometer output signal during the period when the experimental runs were made (see also Appendix B). It was unfortunate that the source of this drifting could not be located and removed, because the requirement for a good product concentration measurement is essential if any feedback concentration control system is to meet with total success. If the refractometers cannot be depended upon to hold a calibration curve--after they have undergone a complete checkover and are within their performance specifications--it will be necessary to develop or buy a more dependable in-line concentration measuring device.

(b) Noise in Process Measurments

Andre [(2), p. 109] encountered quite severe noise problems in the liquid level signals and in the major flow signals transmitted from the unit. He attributed the noise in the level signals to the boiling and frothing in the first effect and to surging of the solution through the second effect. These pressure disturbances could then be passed from the measured variables through the controllers to the controlled variables, where they were once again detected as oscillations on the chart. Pen dampers had been added to many of the recorders to make the readings legible.

To overcome this noise problem, and at the same time provide a visual indication of the true liquid level in the first effect and in the separator, external level indicators constructed from a four foot section of 1.5 inch lucite tubing were connected to the first effect and

the separator. A drawing of the first effect level indicator and its instrumentation is presented in Figure 8. The liquid level in each indicator was then transmitted to the respective Foxboro level controllers. The level indicators are, in effect, first order systems with adjustable time constants. The capacity of each indicator is constant, but the resistance to fluid flow from the process to the indicator can be changed by restricting the inlet flow area (valves V46 and V48). The time constant for each indicator was determined as the time at which 63.2% of the indicator response was complete, following a step change to the differential level between the process liquid and the indicator liquid. The average time constant with the inlet valve wide open for the first effect level guage was 0.48 minutes and for the separator level indicator 0.50 minutes. The liquid inlet valves were left wide open during all the experimental runs. Any closure of these valves would increase the time constant and the damping action of the level indicators. The 0.5 minute indicator time constant is only 1/20 of the approximate time constant for the holdup in each effect. Disturbances in the actual liquid levels are therefore for all practical purposes, followed almost immediately by the level indicators.

The pen dampers were removed from all recording pens after the level guages were installed. As can be seen from the typical records in Appendix E, the liquid level recordings, and generally all the process recordings, were very smooth during the experimental runs. There was still some continuous oscillation in the flow recordings of the first effect bottoms and product flows. However, since the product flow recording was completely smooth when a new valve with a built-in valve

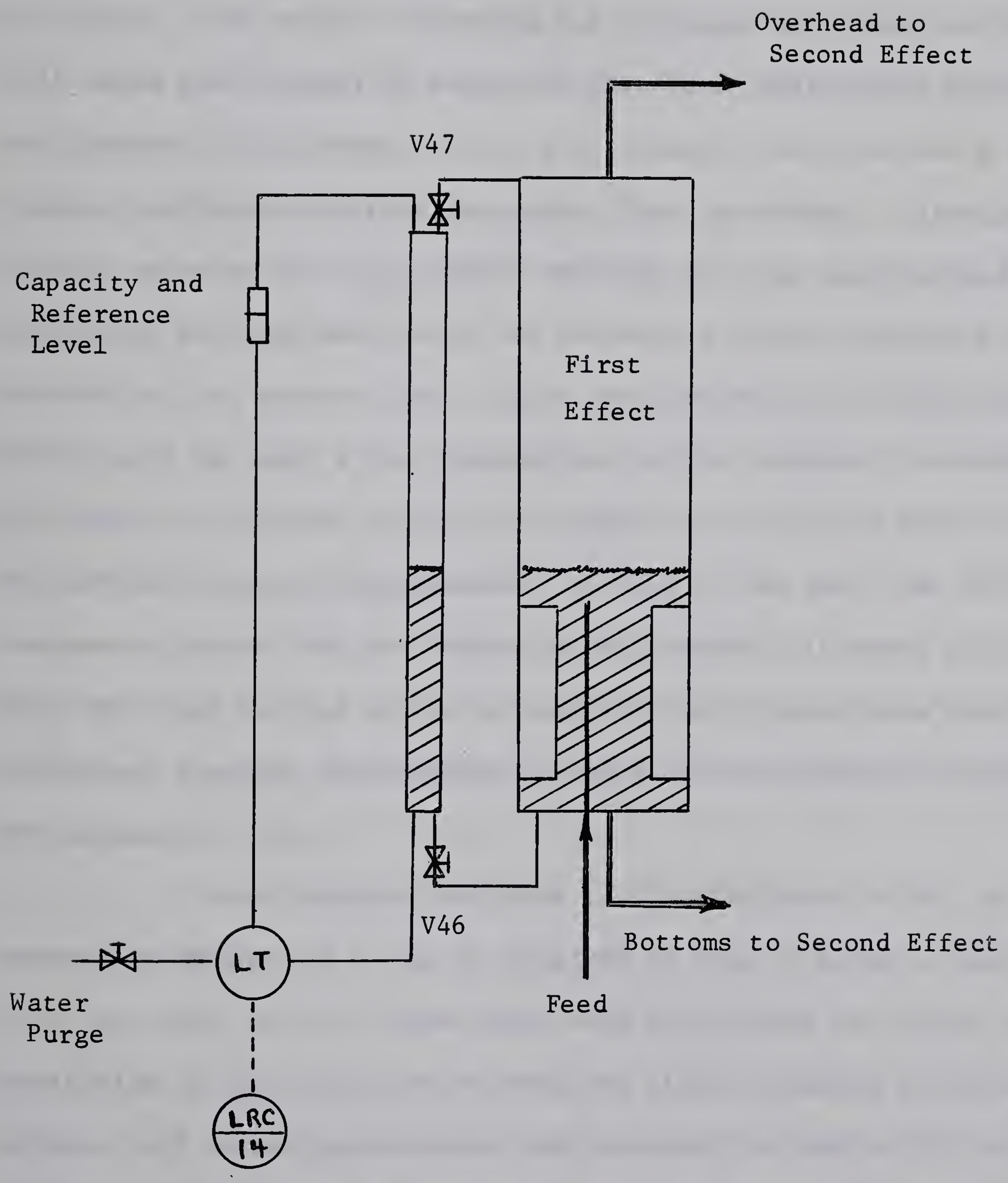


Figure 8 FIRST EFFECT LEVEL INDICATOR AND INSTRUMENTATION*

*Separator level indicator installation is identical.

[illegible]

[illegible]

positioner was first installed, it is logical to assume that the oscillations were primarily the result of a sticky sugar coating on the control valve components which tended to retard smooth operation of the valve. Both valves contacting the hot sugar solutions are equipped with valve positioners; it was hoped that these positioners would hold oscillations to a minimum. It is also evident that there was a small regular oscillation in both condensate flow recordings. Although the control valve on the first effect overhead line was serviced and various controller settings were tried for the second effect condensate level controller, the cause of this slight oscillation was not discovered. Since there was only a flow transmitter in the condenser condensate line, the slight oscillation in this flow signal can only have been the result of upstream pressure disturbances. In view of the fact that both condensate signals are not damped at the recorder, it seems quite plausible that the rough boiling action in the two effects might have developed sufficient pressure disturbances to cause the observed oscillations in the condensate flows.

In no instance are there sharp noise peaks in any of the records in Appendix E as can be observed on some of Andre's recordings [(2), pp. 4-14, 4-15]. These peaks were very likely the result of cavitation in the recirculation pump and liquid slugging in the second effect. All of the experimental runs reported in Chapter VII and Appendix G, with the exception of Run No. 10L, had a steady flow of solution through the second effect. During Run No. 10L the second effect pressure was dropped to its minimum value of 3.0 psia. Cavitation in the recirculation pump resulted, and sharp flow disturbances were

transmitted from the recirculation flow signal to the first effect bottoms and the condenser flow signals. It therefore appears that the installation of the centrifugal pump and the larger recirculation line has not only improved the heat transfer in the second effect, but has also helped to decrease the noise content of many of the process variables.

B. Evaporator Operation

1. Start-up Procedure

The problems associated with the evaporator start-up have been explained previously [(2), pp. 73, 74] and are inherent in the evaporator operation during any start-up. Rather than repeat a discussion of these problems, it was felt that a detailed outline of the evaporator start up procedure, including a working combination of controller set points, would be of value. Such a procedure is included in Section B of Appendix A.

Some points concerning the start-up procedure are worthy of special mention so that the start-up procedure may be more easily followed.

(a) The dp cells must be bled before the system is evacuated.

(b) The product pumps must be running before the system is evacuated to prevent the back flow of air from tankage through the pumps to the evacuated process.

(c) The evaporator system should be evacuated before steam is supplied to the first effect calandria to prevent extreme flashing and carry over of the heated solution once the pressure falls sufficiently to permit boiling.

(d) Liquid levels should be moderately low during start-up to prevent solution splash over when boiling begins.

(e) The product concentration response is slow, but can be hurried somewhat by setting the steam flow approximately 20% too high during initial start-up, and reducing it to the desired value when product concentration approaches a steady state value.

(f) A purge bypass line is provided to assist in the removal of non-condensables from the first effect and second effect steam chest during start-up. The capacity of this bypass is much too large for normal operation: it therefore must be closed once steam begins to flow through it.

(g) Manual control for the liquid levels is necessary during start-up. Once levels stabilize after boiling has begun, the level controllers may be switched on automatic.

2. Normal Operation

The problems encountered during normal operation of the evaporator were not nearly as numerous or as serious as those which make the start-up procedure so complicated. The majority of these problems have already been discussed in Section A of this chapter or by Andre [(2), pp. 74, 75].

The evaporator feed system was the major bottleneck which remained to prevent smooth, continuous operation of the evaporator. Feed had to be prepared batch-wise in each of the two feed tanks in quantities sufficient to maintain approximately seven hours of evaporator operation. There was absolutely no control over the concentration and temperature of the feed, other than the judgement of the operator in determining how much water and product from the previous run should be mixed together to form the next batch of feed. As feed was being taken from and

recirculated through one tank by the feed pump, the new feed had to be remixed in the other feed tank by the auxilliary pump. Even though the feed was normally mixed for approximately four hours, it was not mixed perfectly; feed concentration changes during an experimental run were observed for a number of the runs reported in Appendix E. The feed and auxilliary pumps were Fostoria Dynapumps, which are cooled by giving up heat to the solutions which pass through them. Therefore, in addition to the increase in temperature of the feed solution as a run progressed (this effect became most serious when the tank holdup was small), there was also a steady temperature buildup in the new feed being prepared. This, when combined with the effect of an imperfectly mixed feed, often resulted in a noticeable change in feed properties during an experimental run.

The experiments were planned in such a manner that two runs would be made using the same feed tank. The first run was, of necessity, a feed composition change--the second could involve either a feed flow rate change or a product concentration set point change. Since each run required approximately three hours to complete, only one hour was left to set up the second run, balance the feedforward controller (see Appendix C), take the initial and final steady state readings for the last run, and then prepare for the switching of feed tanks. Of necessity, the feed tanks were often run almost dry before the second run could be completed.

In view of the difficulties encountered with the present feed system, it is recommended that a continuous feed system be designed for the double effect evaporator. This system should be able to continuously remix the concentrated sugar product with water from the main condensate

tank, supplying feed at a specified temperature and concentration. Such a feed system would also remove the need for temperature compensation in the feed concentration measuring refractometer since it could be installed in a constant temperature portion of the feed line.

CHAPTER V

EVAPORATOR PROCESS CONTROL AND INSTRUMENTATION

A. Choice of a Control System

At the time this project was begun, the evaporator had already been used to test a number of different control systems [(2), pp. 61-67]. Three of the four systems had been unstable, but the fourth was successful and considered by Andre as the "best" control system to provide product concentration control for the double effect evaporator. It involved regulation of the steam flow via the product concentration controller. Level control was achieved by regulation of the solution flows from each effect. The complete control system is outlined in Figure 13 of Andre's thesis and is presented in this report in Appendix A with the modifications already discussed in Chapter IV.

Any evaporating process can be viewed very simply as one which converts a dilute feed stream into a concentrated product stream and a number of condensed vapor streams. Cooling water and steam are the other process inputs. Figure 9 illustrates an evaporator process in this manner and includes the properties of each inlet and outlet stream which may be controlled. Quite reasonably, not all of the variables listed can be controlled simultaneously. Internal variables which also require control, such as liquid levels and operating pressures have not be included in the diagram.

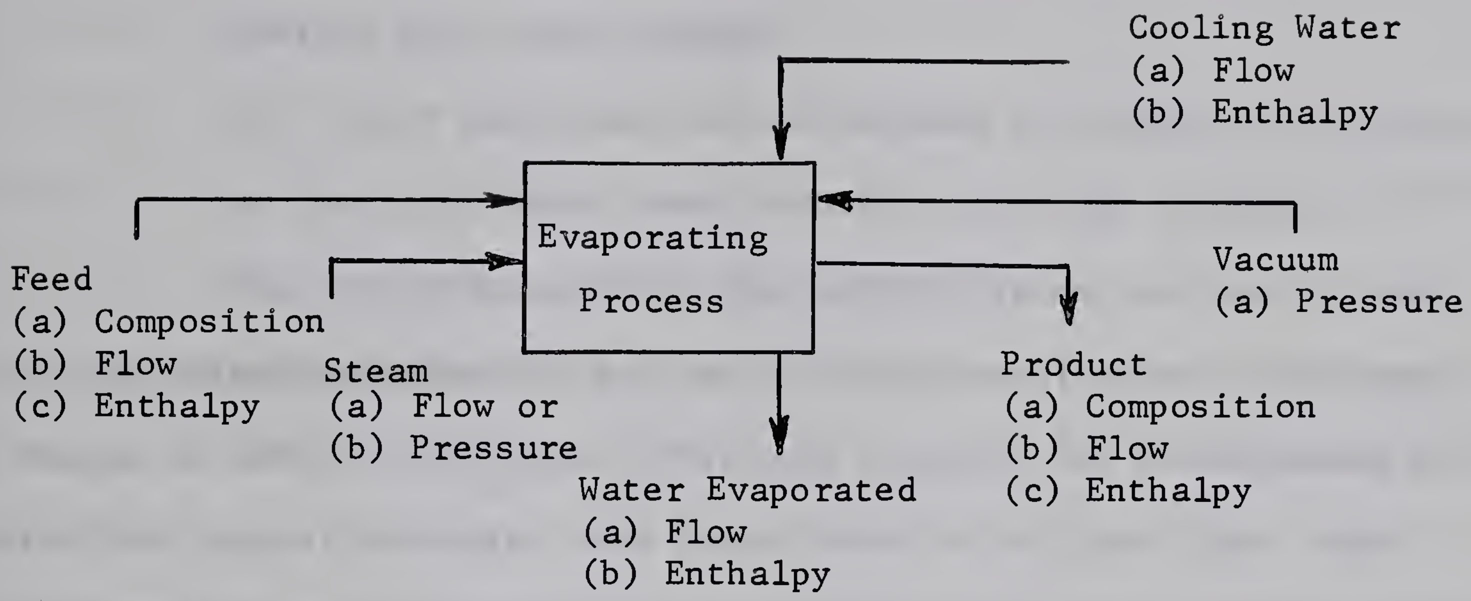


Figure 9 INPUT AND OUTPUT VARIABLES FOR AN EVAPORATING SYSTEM

The control objectives for the double effect laboratory evaporator were:

1. Maintain accurate product concentration control while accepting feed of random composition at a random feed rate.

2. Provide material balance control for the evaporator so that

   (a) the average flow of product and pure condensate streams from the unit will always equal the average flow of fresh feed.

   (b) adjustments to process streams are made smoothly following disturbances to the system.

3. Maintain the process variables within certain predetermined constraints at all times:

   (a) all process liquid levels must be confined within the safe operating limits of the system,

   (b) the vacuum source and pressure control system must be able to maintain a constant operating pressure in the last effect. This pressure should be the minimum economically

available (19) to provide maximum steam economy,

(c) the recirculation rate in the FCE must be adequate to sustain good heat transfer,

(d) inert gases must not be allowed to collect in the system to the point where they interfere with heat transfer.

The disturbances which the control system may have to cope with may enter the system in any one of the process inputs or through a change in ambient conditions. For this project, the disturbances for which the control system(s) were tested were in the feed flow, feed concentration, or in the product concentration set point. Every attempt was made to hold all load conditions (other than in the disturbed variable) constant during all experimental runs.

The above specifications on the design and operation of a suitable concentration control system are very similar to those which were inherent to Andre's study. Therefore, Andre's "best" control scheme appeared to be the logical choice for the basic evaporator control system.

The feedforward product concentration controller then had to be designed and interfaced with the Foxboro electronic instrumentation on the evaporator. Since the feedforward controller determines the steady state steam flow required at any set of operating conditions (see Chapter II), it appeared that it could be readily integrated into a control system if the steam flow controller could accept a cascaded set point. The control system which existed on the evaporator was ideal for adaptation to feedforward control: instead of the steam flow set point being determined by the Foxboro concentration controller, it

would be set by the feedforward controller. When feedback trimming of the feedforward steam flow setting was desired, the corrective action calculated by the Foxboro concentration controller could be added to the steam flow set point signal from the feedforward controller. The remainder of the evaporator control system was unaffected by the integration and functioned as it had during previous experiments.

B. Modifications to Evaporator Instrumentation

Once the basic evaporator control scheme had been determined as suitable for integration with the feedforward concentration control system, it was examined closely for areas of possible improvement. The evaporator modifications which involved changes in the instrumentation are summarized in Table 3. The reasons for these modifications have already been fully discussed in Chapter IV.

Table 3

CHANGES TO PROCESS INSTRUMENTATION

Instrumentation Change	Effect
1. Control valve removed from recirculation line in second effect, large orifice run installed in new 1.5 inch circulation line	Increased recirculation flow permitted upon installation of new pump
2. Automatic level control system removed from condenser. Spillover level control system installed	Condenser flooding eliminated at all operating pressures
3. Level transmitter taps removed from first effect and separator and placed on level indicators	Noise content of transmitted level signals was greatly reduced
4. Condenser pressure control valve relocated in air bleed line to condenser	Manual loading of condenser vacuum system not required

A further change in the method of obtaining condensate temperature and pressure control in the condenser was attempted but did not meet with success. The condenser control scheme of Figure 10.1 was used by Andre. In order to provide stable pressure control, it had been necessary to omit condenser condensate temperature control, and operate with the cooling water manually set at a flow rate sufficient to ensure total condensation of all the vapor entering the condenser. It was also necessary to manually adjust the atmospheric intake valve, so that the pressure control system would be loaded sufficiently to prevent on-off pressure control. To make the pressure control system completely automatic and to avoid the need for throttling the vacuum supply to the condenser, the vacuum control valve was installed in the air intake line. At the same time the condenser control scheme was switched to that of Figure 10.2 in the hope that close pressure control could be obtained using a control scheme which would minimize the consumption of cooling water. Buckley [(3), p. 265] recommends this "thermo-gas release" system because it "avoids the instabilities and other control difficulties that often characterize condensate temperature control systems based on the manipulation of the condenser cooling water." The difficulties he mentioned as inherent to the control of condensate temperature via cooling water manipulation (Figure 10.1) are:

1. The condenser control system tends to become unstable when two or three mode controllers are used for the temperature controller.
2. Cooling water flow is either insufficient or excessive.

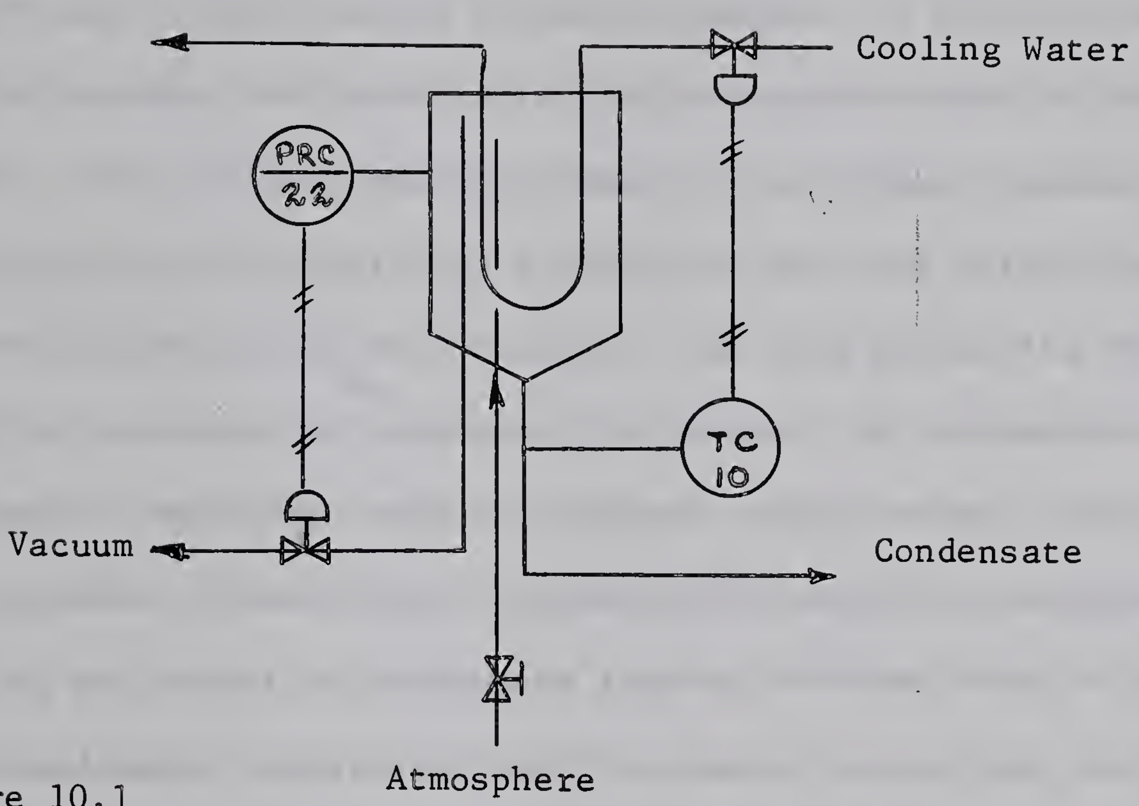


Figure 10.1
Condenser Controls - Andre (2).

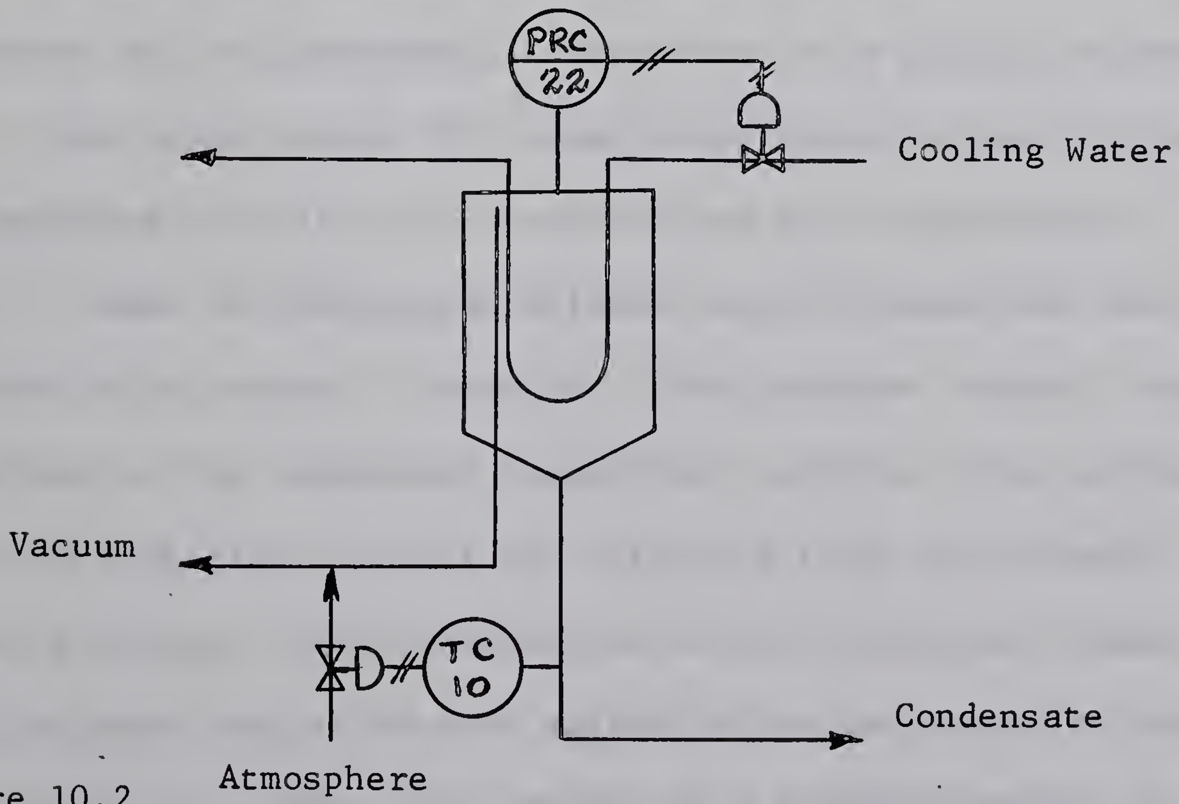


Figure 10.2
Condenser Controls - Buckley (3).

Figure 10 CONDENSER TEMPERATURE AND PRESSURE CONTROL*

*Auxilliary instrumentation has been omitted for clarity.

3. The temperature control system must overcome the slow condenser heat transfer dynamics.

Both Buckley (3) and Pyle (16) endorse the thermo-gas release system for use on distillation column condensers. A distinction between the control systems for distillation and evaporation must be made at this point. For the evaporation process, close product control can be obtained only when the operating pressure in the last effect (and therefore in the condenser) is held constant. As long as all the vapor entering the condenser is condensed, the control of condensate temperature is unnecessary and serves only to conserve cooling water. For distillation processes, close product concentration control is achieved by controlling the amount of condensate that is refluxed back to the column. Thus the condensate temperature must be closely controlled, while the actual condensing pressure is of less consequence in determining the product concentration. There is also a definite relationship between the pressure and the condensing temperature for a single component system. This relationship will cause interaction between the pressure and temperature control loops when both are set independently.

When the thermo-gas release control system was used on the evaporator in an attempt to maintain close pressure control, unstable oscillations in the condensate temperature resulted. The pressure control was essentially on-off and permitted large disturbances in the condenser pressure. Close examination of the thermo-gas release control system indicates that it is best suited to provide close tail-pipe temperature control and that an averaging type pressure control is required if the control system is to work favorably [(3), p. 269]. It therefore

cannot be applied successfully to an evaporating system where very close condenser pressure control is required.

The condenser control system was therefore returned to that of Figure 10.1, with the exception that the pressure control valve remained on the air intake line and the vacuum line to the condenser was left unrestricted. The temperature controller was left on manual operation for all runs, with the water flow set at a rate such that approximately 7% of the total condenser heat load was due to cooling of the condensate below its condensing temperature. The pressure control was very satisfactory as the actual chart records in Appendix E indicate. The return to this type of condenser control once again resulted in excessive cooling water consumption, but this was more than compensated for by the excellent pressure control obtained.

C. Feedforward Product Concentration Controller

The very nature of a multiple effect evaporation process makes it difficult to control. The relatively large solution holdups and the resistances to the flow of mass and heat through the unit, combine to produce relatively large mass and heat transfer time constants in each effect. Feedforward and/or ratio control can be used to advantage for controlling a process such as a multiple effect evaporator, where the time constants and transportation lags associated with each effect are all combined in series.

The major characteristic of a pure feedforward control system is that it uses only system inputs (variables or set points) and a model of the system, to compute the setting (s) for other process variable(s). If the input signals are correct and the model is accurate, each setting

will hold its appropriate process variable on specification, during and after any system disturbance that the model is capable of detecting. Since the control system measures system inputs, the measurements are not delayed or distorted by the process dynamics and compensation for disturbances can be programmed to reach the process simultaneously with the actual disturbances. However, there is a negative aspect of feedforward control systems to be considered also. If any of the process measurements are incorrect or if the model is in error and does not represent the actual system accurately, the feedforward controller output, and therefore the controlled variable, will be in error. Since a pure feedforward system receives no indication of the errors which may exist in its controlled variables, it cannot eliminate them. It is for this reason that some feedback control is usually included in a feedforward control system. This feedback "trims" the controller output, when necessary, to prevent a continuous offset in any of the controlled variables.

The model developed in this study for the double effect evaporator was based primarily on a simple, steady state material balance (see Chapter II). The model relates the system inputs, feed flow and concentration and product concentration set point, to the steady state steam flow requirements using the relationship:

$$S_{MB} = W_o K_s (1 - C_o / C_{psp}). \qquad \text{(II-7)}$$

Further steam flows to supply latent heat to the feed (S_{LH}) and to offset heat losses (S_{HL}) are also computed within the controller. The resulting total steam flow requirements are $S = S_{MB} + S_{LH} + S_{HL}$. (II-9)

It is logical that steam flow rate should be choosen as the variable which would be manipulated to maintain the product concentration

on specification, since it is the major energy input to the evaporating system. (Another possibility might be the pressure in the last effect, but the model linking the process inputs to this pressure would be considerably more complicated than the model developed in Chapter II. Also,the highest steam economy is obtained when this pressure is held to a minimum.) It was fortunate that the previous evaporator control system controlled product concentration via regulation of the steam flow [(2), p. 68]. The evaporator could therefore be switched from feedback to feedforward product concentration control by simply supplying the cascaded steam flow set point from the feedforward controller, instead of the Foxboro concentration controller. It has been noted previously that the product concentration response was quite sensitive to changes in steam flow. This characteristic of the evaporator control system is not essential for the successful operation of a feedforward controller; it is in fact a slight handicap, since this sensitivity tends to magnify the effects of slight errors within the feedforward controller.

A first order system having a ten minute time constant was added to the feed concentration measurement signal in an attempt to simulate the response of the concentration in the first effect (which is essentially a first order system) following a change in the feed concentration. Since there was approximately a minute of dead time before the feed reached the first effect after passing the feed refractometer, a corresponding lag was manually added to the transmitted concentration signal. This was very easy to do since the feed concentration was reset on a potentiometer within the feedforward controller following a change to the feed concentration. The response of the model to feed flow rate and

product concentration set point changes was immediate, and resulted in a rapid step change of the steam flow to its new steady state value. The controller model remained very simple in comparison to the dynamic model of the evaporator developed by Andre [(2), pp. 16-47]. It was not expected that it would be able to hold the product concentration on specification during transients, and it was also anticipated that errors in the model might produce some offset in the product concentration under steady state conditions. Therefore provisions were also made to add negative feedback control to the controller, so that these steady state offsets could be eliminated.

The model is based on a steady state material balance around the evaporating unit and is not capable of supplying extensive dynamic compensation for disturbances during the transient periods which follow the disturbances. The gain of the controller must be one since it computes the exact steady state steam flow required for a given set of operating conditions. In a recent article (10), King, Rothfus and Kermode present a method of solving the dynamic equations describing a distillation column to determine the perfect feedforward controller settings required to prevent disturbances in feed rate and composition from affecting the product concentration. This technique is beyond the scope of this study, but may certainly be of great use when a feedforward control system is designed for the evaporator using a dynamic model such as the one developed by Andre.

1. Controller Components

The model developed in Chapter II and discussed above was programmed on a specially built analog computer. The analog circuit

required to generate the set point for the steam flow controller is shown in Figure 11. The symbols used are explained in Appendix C. The analog computer was constructed from individual function modules which are manufactured by Consolidated Electrodynamics Corporation. Each module will accept signals of 0 to ±10 volts and will generate an output signal which may be the sum, product, ratio or some other predetermined function of the inputs. The specified accuracy of the output varies for module types but 1% is generally the maximum error. For example, the multiply-divide unit has a maximum error of 0.1%; the adjustable coefficient may have calibration and linearity errors of 0.3%. The gain of each module is determined by the resistances of its input and output resistors. Input resistors may be combined to yield gain selections from 0.1 to 15.0; output resistors from 0.5 to 8.5. The tolerance specified for these gain settings is ±1.0%. It is evident that an unfavorable combination of resistor and module errors could produce a significant error in the computed module output.

A check on the performance of the computing system is summarized in Appendix D. Adjustable coefficients and function segments were accurately set to within 0.001 volts using a 10.000 volt reference and a digital voltmeter. The error in these units was therefore negligible. The largest error--0.093 volts at add-subtract unit No. 4--was the result of a low precision (5%) input gain resistor. This resistor was installed on the bias signal input so that its error would be automatically cancelled out by the controller balancing procedure outlined in Appendix C.

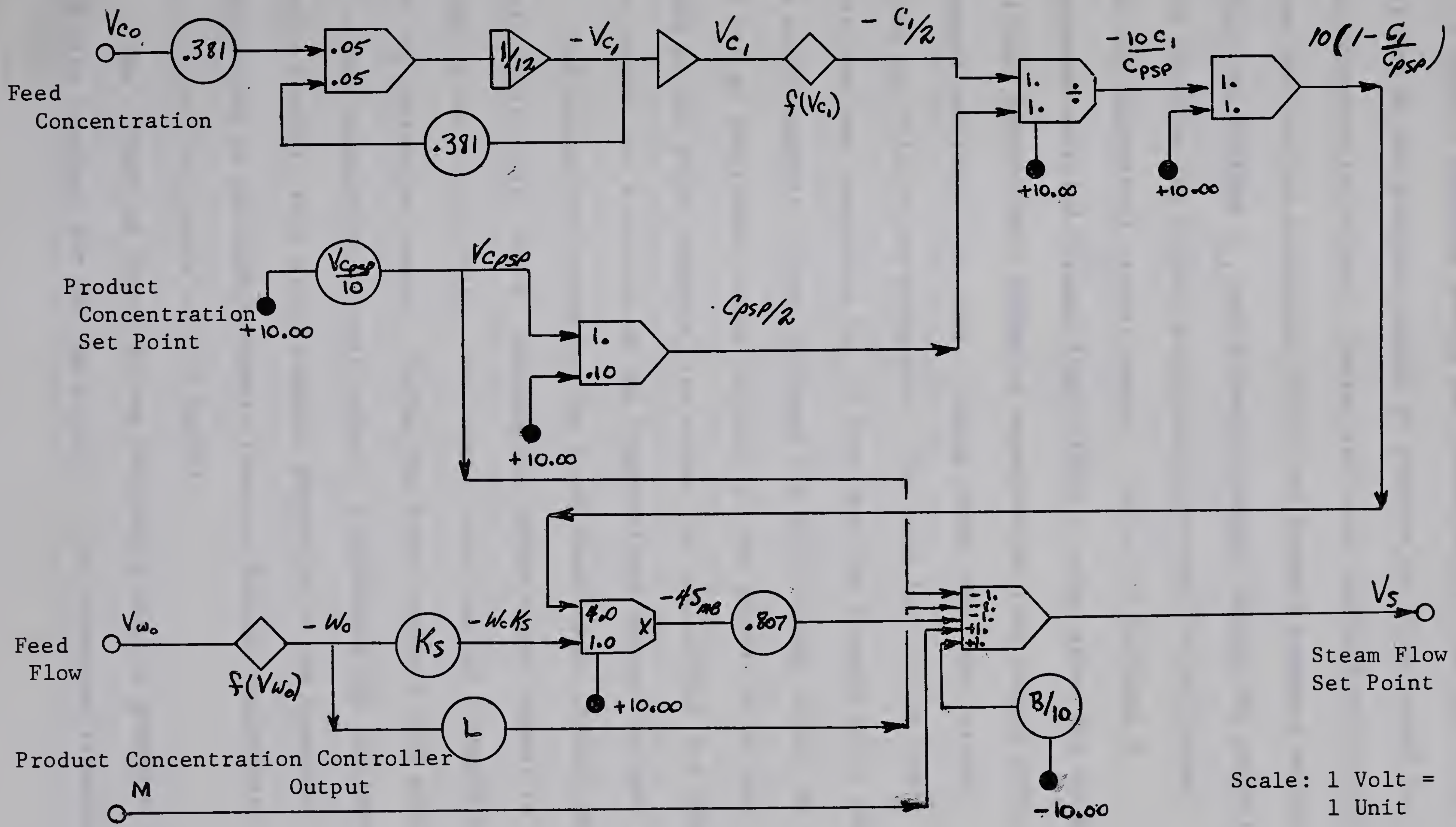


Figure 11 SIMPLIFIED CIRCUIT DIAGRAM FOR FEEDFORWARD PRODUCT CONCENTRATION CONTROLLER

A dynamic response module was connected with two adjustable coefficients and an add-subtract module to provide the first order lag in the feed concentration signal. During the error analysis the initial condition was held on the integrator within the dynamic response module to insure that the voltage V_{c1} would remain constant during the analysis. Consequently the time lag circuit error was quite small--0.141 volts or 1.73% of the theoretical circuit output. Table D-1 includes a column of numbers entitled "Steady State Offset," which gives the actual time lag circuit errors for a number of experimental runs during which the feedforward controller was used. During these runs, the initial condition was not held on the integrator. The time lag circuit error during these runs was comparable to, or less than the circuit error noted during the error analysis. Thus, although the grid voltage offset at the integrator was an additional source of error to the circuit which was not included in the error analysis, its effect was not serious and indeed appeared to cancel out a portion of the computational errors for some runs.

The operational error for the total feedforward circuit was a reasonably small -0.105 volts. Because the system output signal is in the range of one to five volts, the circuit error may be a significant percent of the controller output. During the error analysis this error was 6.70% of the theoretical circuit output. Considering the complexity of the analog circuit, this error compares favorably to the tolerance which is permitted in the output signals from the Foxboro controllers on the unit (1% of full scale or 0.10 volts).

The circuit of Figure 12 was originally used to generate the ten minute time constant for the lag circuit. The resulting transient

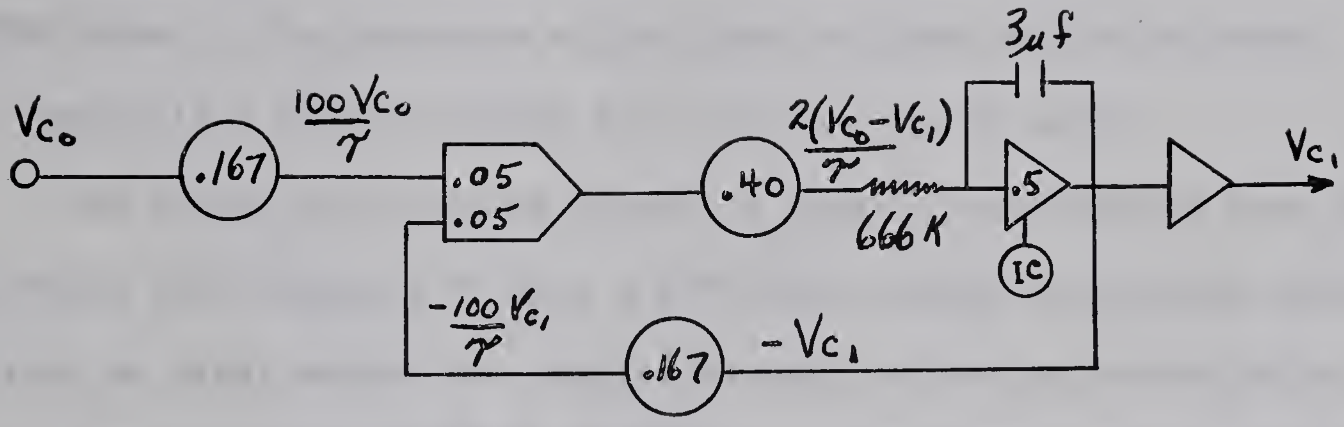

V_{c_0}
.167
$\frac{100 V_{c_0}}{\tau}$
.05
.05
.40
$\frac{2(V_{c_0}-V_{c_1})}{\tau}$
666 K
3μf
.5
IC
V_{c_1}
$-\frac{100 V_{c_1}}{\tau}$
.167
$-V_{c_1}$

Figure 12   INITIAL TIME LAG CIRCUIT

response of the circuit following a step change in V_{co} was 63.2% complete in ten minutes, but the steady state output concentration V_{c1} was often offset from V_{co} by as much as two volts. This offset was not always in the same direction. This behavior originally caused much concern about the ability of the modules to produce large time constants. It was however, noted from the dynamic response module specifications that an input grid voltage offset of up to 0.010 volts may be encountered. The circuit of Figure 12, which was to integrate the signal $2(V_{co} - V_{c1})/T$ to zero, could therefore stop integrating once the input signal was within the range $[-0.010 \leq 2(V_{co} - V_{c1})/T \leq +0.010]$. The reason for the large offset in the delayed concentration signal was now quite clear. Even with a maximum input of $V_{co} = 10.0$, the small loop gain cuts the grid voltage to 0.033 volts. The effect of a 0.010 volt offset would therefore be to produce a 30% error in the output voltage.

It was apparent that the input grid voltage must be kept as large as was possible, if the effect of the grid voltage offset was to be minimized. This required that much more of the loop gain be transferred

to the integrator. This gain was decreased by a factor of 6 by installing a second 3 micro-farad capacitor in parallel with the module capacitor, and by tripling its input resistance (to the maximum permissible value of 2000K-ohms). The magnitude of the input voltage may be increased even further if a smaller system time constant can be used.

The normal grid voltage offset is usually considerably less than 0.010 volts (see Appendix D) with a 0.003 volt offset considered typical. Therefore we might expect the revised circuit to have a concentration offset of approximately 1/18 of the 30% error mentioned above. The steady state concentration offsets listed in Table D-1 for a number of experimental runs support this supposition.

2. Interfacing the Feedforward Controller with Evaporator Instrumentation

All of the recorded and controlled variables on the evaporator are transmitted by the electronic Foxboro instrumentation as 10 to 50 milliampere signals. Precision 1%, 200 ohm resistors were installed in all the major control loop lines, so that the 2 to 10 volts potential drop across the resistor could be transmitted to a multi-variable, differential voltage recorder. Theoretically, any of these signals could be transferred to the feedforward concentration controller in this manner.

Figure C-1 illustrates the actual hookup by which the inputs were to be brought into the feedforward controller. In practice, feed temperature variations and the resulting indicated concentration variations could not be avoided if the feed refractometer was to be used without temperature compensation. Therefore the feed was analyzed on the bench refractometer before each run, and the voltage corresponding to the

correct feed concentration was set on a potentiometer within the feedforward controller. This procedure permitted the controller to operate as if all its inputs were coming directly from the evaporator instrumentation as in Figure C-1, but removed one source of error from the input variables.

To permit the transfer of signals from the feedforward analog controller to the Foxboro control loops, some further modification to the controllers was necessary. The Foxboro controllers normally drop both the cascade set point signal and the measurement signal (10-50 m.a.) across a 0.1% 100 ohm resistor to generate voltage representations of these two signals. These voltages are then used by the controller to compute the error in the controlled variable. Therefore by removing the 100 ohm resistors, the set point or measured signal to the Foxboro controller can be respresented by a 1 to 5 volt input. It was essential that the reference potential for all voltages transmitted between the Foxboro instrumentation and the analog controller be identical. The manner by which a common ground was obtained for all the evaporator instrumentation is illustrated and discussed in Section B of Appendix C.

It was necessary to program into the feedforward controller a conversion equation for each of the four voltage signals which were transferred to or from Foxboro instrumentation. After conversion, the voltages representing W_o, C_o, and C_{psp} were the voltage equivalent of the physical values of these variables. The computed steam flow in volts (representing pounds per minute) is then converted to a voltage set point signal and transferred to the steam flow controller. Due to the shortage of analog modules, each calibration curve had to be

represented by a linear relationship. As can be seen in Appendix B, a straight line represents both concentration calibration curves very well, but is successful in representing the feed and steam flows accurately for only a limited range of the controller output (or input). Table B-1 in Appendix B summaries these linearized calibration equations and gives the voltage range within which the conversion equation error is less than 5% of the correct value. Since both flow conversion equations are parabolic, the error in the linear approximation to these curves increases exponentially as the transmitter outputs fall below the recommended lower operating extremes. For this reason, feed and steam flows were held closer to the upper calibration extremes than to the lower extremes during the experimental program.

The functions described by the conversion equations in Table B-1 were generated using the function segments and add-subtract units as represented in Figure 11.

It would have been desirable to bring each voltage input signal into the computer through a high input impedance amplifier. It was necessary, however, due to the shortage of computing modules, to bring some inputs in through 10K-ohm input impedances, such as adjustable coefficients and function segments. Thus the computing circuit may consume up to 1 milliampere when the input is 10 volts. This means that an error of approximately 2% could result when the signals are transferred from the Foxboro control circuits (in which the current is 10 to 50 m.a.) to the analog controller. It is possible to lower this error to .2% if all inputs can be introduced through a 100K input impedance, such as an add-subtract unit with an input gain of 5.0. Two hundred K-ohm is the

maximum input impedance available on any C.E.C. module; it was not possible however, to rearrange the modules in the circuit of Figure 11 so that input impedances could be increased above their 10 K-ohm values.

3. The Nature of Feedback Compensation in the Feedforward Product Concentration Controller

As mentioned previously, some feedback compensation to the feedforward steam flow signal was included as an option in the analog controller (see Figure C-4 for a block diagram). There were two alternatives for the provision of this control action; the measured product concentration could have been transferred to the analog controller where the concentration error and the corrective action would be computed, or the corrective action as determined by the Foxboro concentration controller could be added to the analog controller output. The latter alternative was choosen because:

(a) the extra modules required to construct a two or three mode controller within the feedforward unit were not available,

(b) even with adequate computational capacity, an analog 2 or 3 mode feedback controller would be less flexible and more difficult to adjust than the Foxboro controller when large time constants (approximately 30 minutes) are to be generated,

(c) feedback compensation for the feedforward controller is readily available but is not built into the controller.

As a consequence of (c), the evaporator was easily switched to open loop, feedforward, or feedforward-feedback product concentration control without rewiring *any* of the evaporator instrumentation. In practice it was difficult to transfer the evaporator to straight feedback control only, since the automatic balancing procedure on the Foxboro controller was

lost when it was modified to receive a cascade voltage set point. In this instance, it would be advantageous to remove the cascade set point and thereby recover the balancing action when a significant number of conventional feedback experiments are required.

The switching procedure by which each of the types of concentration control was obtained is fully explained in Appendix C. The manner in which the corrective portion of the Foxboro concentration controller output was added to the analog controller output is also illustrated in Figures C-1 and C-2.

The Foxboro concentration controller current output was converted to a 1 to 5 volt signal by taking the potential above a grounded 100 ohm resistor as the output. At all times this voltage consisted of the sum of the product concentration set point and the feedback correction voltages. By simply subtracting the set point voltage from the net concentration controller output at the last add-subtract unit (AS-4), the feedback correction signal computed by the Foxboro controller could be added to the feedforward controller output.

When the evaporator was to be run under feedforward concentration control only, the Foxboro controller output was switched to become the measured variable for the controller, and the controller action was increased to approximately 100% proportional band and 0.20 minutes per repeat. This procedure balanced the controller output against the product concentration set point so that, at all times, the net feedback compensation was zero. To eliminate feedforward control, the Foxboro concentration controller output was switched to bypass the feedforward unit and was cascaded directly to the steam flow controller set point.

Further details describing the operation of the feedforward controller using its different control options are given in Appendix C and will not be repeated here. A detailed balancing procedure by which the analog controller was set to provide a bumpless transfer of the steam flow set point from one control system to another is also presented in that appendix.

4. Possible Alternate Use for Evaporator Feedforward Control

At the time the feedforward system was developed for use in conjunction with the previous evaporator control scheme, possible alternate evaporator control systems, into which the feedforward controller could be integrated, were also considered. Figure 13.1 is a schematic diagram of the control system which was used during this study. Only the basic control loops and instrumentation are represented on the diagram.

Because the discharge from each effect is regulated by the level controllers on the respective effects, the control system of Figure 13.1 will not allow flooding of either effect during start-up or following a disturbance. However, off-specification product can be produced by this control system during periods when the liquid levels become too high. To prevent large amounts of off-specification product from being produced during start-up, the feed rate was manually set quite low and the steam rate quite high, so that steady state conditions can be reached in a minimum amount of time. Andre reported a very quick steam pressure (and therefore heat transfer) response from the first effect calandria to the first effect vapor space and onward to the second effect steam chest and vapor space [(2), p. 66]. This characteristic is useful when feedback product concentration control is being used, but is

Experimental Control System
(Transmitters and some auxilliary equipment
are omitted)

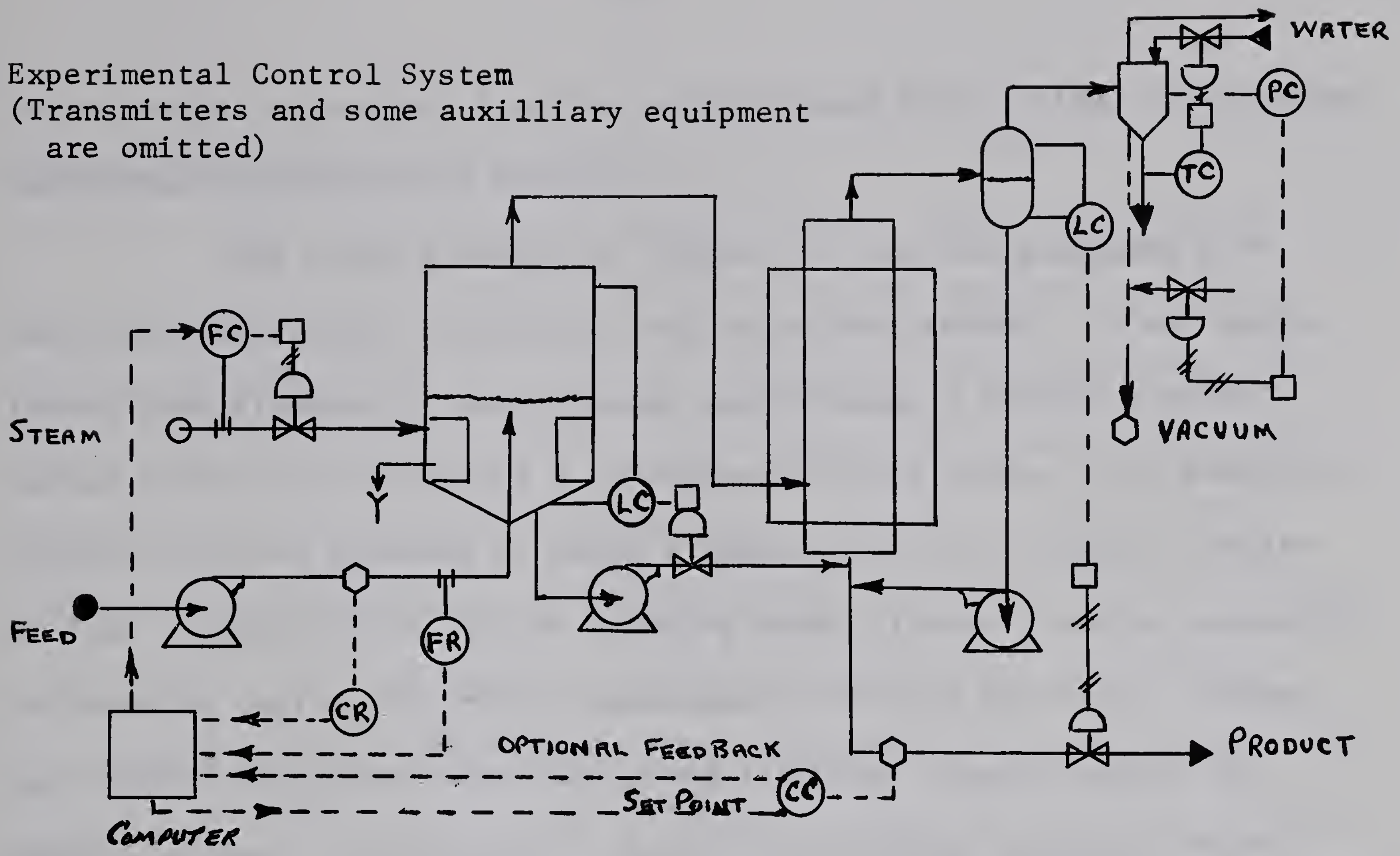


Figure 13.1

Possible Alternate Control System
(Condenser control is unchanged from Figure 13.1)

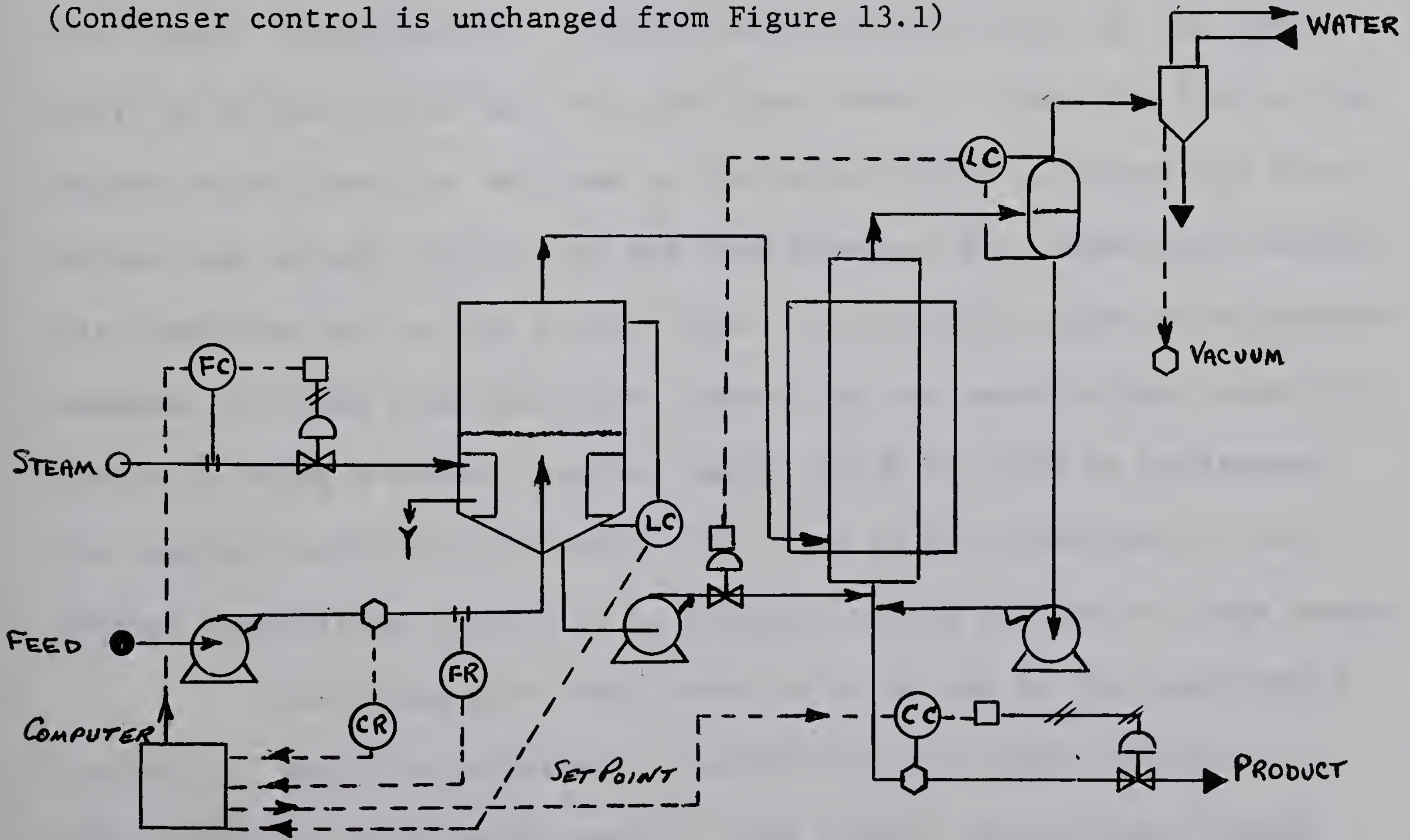


Figure 13.2

Figure 13 CONTROL SYSTEMS FOR THE DOUBLE EFFECT EVAPORATOR

not essential to prevent flooding in the system (or to allow satisfactory feedforward concentration control).

The control scheme of Figure 13.2 was the alternate with which the feedforward controller could have been tested. It was anticipated that flooding in both effects could become a serious problem during start-up or following a disturbance to the system. The start-up problem could be overcome by using a manual start-up procedure similar to that outlined above, and by allowing some off-specification product to be produced during the period immediately following start-up. However, any sizable disturbance may also cause flooding, simply because the control system is attempting to satisfy two somewhat opposing control objectives. It accepts feed at a random rate while trying to maintain strict concentration control on the product. If the feed rate were high during a period when the product was being throttled to build up the product concentration, flooding would quickly result as the level built up in the second and then the first effect. Since the flow to the second effect would be cut down as the level builds up there, the first effect must absorb the bulk of the feed whenever the composition controller throttles back on the product flow. This problem might be alleviated somewhat by using wide band level control on the second effect separator and/or by using a special control system (such as might be programmed for digital control) which would allow feed to be controlled, or the product composition control to be relaxed, during periods of large upsets.

The steam flow rate, which is being set by the feedforward controller, would be adjusted by the first effect level controller to compensate for fluctuating levels. This control system would depend

upon a quick boilup-to-steam flow response, if the first effect level were to be maintained within reasonable limits. These limits must be fairly tight so that good circulation through the calandria can be maintained, and so that the liquid entrainment in the first effect overhead is minimal. Andre found the first effect level quite insensitive to steam flow rate changes. Although the problem he encountered of measuring the actual liquid level of foaming solution has been overcome by the use of external gauges, it is not expected that the response of liquid level to steam flow has been improved by this alteration. It appears that the control scheme of Figure 13.2 may be, at best, difficult to apply using conventional instrumentation. If and when adaptive control systems can be programmed for direct digital control of the evaporator, this "alternate" control system may be able to meet both the control objectives mentioned above, following moderate input disturbances and during normal operation.

CHAPTER VI

EXPERIMENTAL PROGRAM

Once the feedforward product concentration control system had been designed and constructed, it was necessary to determine how well it performed. It appeared that the best manner to obtain this information would be to integrate the feedforward controller with the existing evaporator control system in such a way that the evaporator could be run independently under open loop, feedback, feedforward or a combination of feedforward and feedback concentration control. Similar disturbances could then be introduced into the evaporating system when it was running under each method of product concentration control. The transient response of the product concentration following these disturbances would then be a direct indication of the performance of each control system, and would offer a means by which the control systems might be compared.

It is to be emphasized at this point that the settings for the controllers used in each of the four classifications of experimental runs are not necessarily optimal. These settings were determined prior to the open loop runs by the trial and error procedure which has been described in detail by Haggart (7). A complete list of the controller settings for each run is included with the experimental data in Appendix E. It can be seen that the majority of the controller settings remained the same for all the experimental runs. The changes which were made were usually the result of changing control loop characteristics (for example the percent proportional band was increased from 50 to 150% for the separator level controller for Run No. 12FFB following cleaning of the

product valve). The other major change of controller settings occured before Runs 17FB and 18FB, when the settings on the steam flow and product concentration controllers were returned to the values which Andre had used for his feedback runs. This was done so that the concentration response curves obtained from this study could be directly compared to those presented by Andre, and hence, provide an indication of the effect the changes described in Chapter IV have had on the evaporator performance. The open loop runs from the two reports can be compared for the same purpose.

Since the process variables which were used by the feedforward controller to compute the steam flow set point were feed flow rate, feed concentration and product concentration set point, it was logical that the disturbances for which the control systems were to be tested should be in these variables. For the open loop "set point" disturbance run, the change was actually a direct change in the manual setting on the steam flow controller. This was necessary since there can be no feedback connection between the product concentration and the steam flow rate when the concentration control loop is "open."

The experimental runs reported by Andre [(2), p. 82] involved changes to feed flow rate and product concentration set point (or steam flow rate), while the evaporator was being run either open loop or under feedback concentration control. Some of the disturbances with which these runs were initiated also involved simultaneous changes to the feed temperature and/or concentration.

A summary of the experimental program is presented in Table 4. It can be seen that the planned run involving a set point change in

product concentration, with the evaporator under feedback concentration control, was not done. The previous run involving a feed flow disturbance had resulted in a highly oscillatory concentration response which did not stabilize in 200 minutes; since the open loop runs had indicated that the evaporator was even more sensitive to steam flow upsets than to feed upsets, it was quite certain that the set point change run would be even more oscillatory. The controller settings used by Andre were clearly not suitable for the remodeled evaporator.

Data from all of the runs reported in Table 4 are presented in either Chapter VII or Appendix G. The only experimental runs not reported in this thesis are from a series of feedforward runs which were made before the steady state model represented the actual evaporator performance with sufficient accuracy. For these runs K_s, as defined by equation (II-5), was also choosen to include the system inefficiencies due to heat losses and latent heat requirements of the feed. This definition proved unsuitable since the term K_s was not a constant when it was defined in this manner. Run No. 5FF is included in Appendix G as a representation of these discarded experiments.

Experimental Run No. 10L was a special run in that the disturbance to the evaporating system was in the operating pressure of the second effect. The operating pressure was dropped to the minimum which the vacuum system could sustain (3.0 psia) so that the performance of the new condenser condensate removal system could be observed. This run is also discussed in Appendix G.

One further point of interest with respect to the runs reported in Table 4, is that many of the runs also included a small set

Table 4

EXPERIMENTAL PROGRAM

	Experiment No.[c]	Nature of Experiment
Open Loop:	10L[a]	Open loop response to a step decrease in the operating pressure in the second effect
	20L	Open loop response to a step increase in steam flow rate
	30L	Open loop response to a step decrease in feed concentration
	40L	Open loop response to a step increase in feed flow rate
Feedforward:[b]	5FF[a]	Open loop response to a step increase in feed concentration
	6FF[a]	Open loop response to a step increase in feed concentration
	7FF[a]	Open loop response to a step increase in product concentration set point
	8FF	Open loop response to a step increase in feed concentration (coupled with a decrease in the product concentration set point of 0.20% sugar)
	9FF	Open loop response to a step increase in feed flow rate
	10FF	Open loop response to a step decrease in the product concentration set point (coupled with a starting set point error of +0.08% sugar)
Feedforward & Feedback:	11FFB[a]	Closed loop response to a step increase in feed concentration (coupled with an increase in the product concentration set point of 0.07% sugar)
	12FFB	Closed loop response to a step increase in feed concentration (coupled with an increase in the product concentration set point of 0.07% sugar)

continued

Table 4
(continued)

EXPERIMENTAL PROGRAM

	Experiment No.	Nature of Experiment
	13FFB[a]	Closed loop response to a step increase in feed flow rate
	14FFB[a]	Closed loop response to a step decrease in feed flow rate
	15FFB	Closed loop response to a step increase in feed flow rate
	16FFB	Closed loop response to a step increase in product concentration set point (coupled with a starting set point error of +0.09% sugar)
Feedback:	17FB	Closed loop response to a step decrease in feed concentration
	18FB	Closed loop response to a step increase in feed flow rate (coupled with a decrease in the product concentration set point of 0.02% sugar)

[a] These runs are presented in Appendix G; the remainder in Chapter VII.

[b] The feedforward transient responses have been classified as open loop because there was no information fed back from the Foxboro concentration controller to the steam flow controller during any of these runs.

[c] Experiments were conducted in the order they are numbered.

point change in addition to the major disturbance. This set point change resulted due to the necessity of performing the balancing procedure (Appendix C) and resetting the product concentration set point at the start of each run. Unfortunately, the new set point did not always correspond exactly to the actual product concentration for reasons which are explained in Chapter VII.

In some instances, the disturbances which were introduced to the evaporator system were large enough to cause some of the process variables to go temporarily beyond their calibrated limits. For example, in runs number 40L and 15FFB the first effect bottoms flow exceeded three pounds per minute, which corresponds to 100% of the transmitter output. Similarly in Run No. 16FFB the product valve was indicated as completely closed for a portion of the transient. During this period the flow was calibrated as 0.316 pounds per minute.

Table 5 outlines the scope of the experiments reported here and those discussed by Andre [(2), p. 82 and Appendix 5], and is presented to show that both sets of experiments were run approximately within the same operating region for the major system variables. Since the first effect concentration C_1 was not measured during the above experimental runs, only Andre's figures for this variable are included in Table 5.

It is also worth pointing out that Andre attempted to maintain his feed at approximately 3% sugar by weight for his experimental runs. During the present study, feed concentration was intentionally changed for every second run. In this manner the effect that disturbances in feed concentration had on the product concentration could be determined.

The experimental runs reported in Table 4 were made during four separate three or four day operating periods. In view of the long period that was required following start-up before steady state conditions were reached (approximately six hours), it was considered necessary to take as much data as was possible during a single operating

period. In the periods between experimental runs, the suitability of the data already obtained was checked so that the next series of runs could be planned. In this manner, the operating time consumed during start-ups was kept reasonably small, and no large periods of operating time were used to produce data which ended up being discarded.

Table 5

A COMPARISON OF THE SCOPE OF THE PRESENT EXPERIMENTS TO THOSE OF ANDRE[1]

Parameter	Minimum Value From:		Maximum Value From:	
	Andre	Present Study	Andre	Present Study
W_o	2.43	2.90	3.80	3.87
W_1	1.69	1.60	2.80	3.0 (approximate)
W_p	0.96	0.316	1.84	2.33
C_o	2.78%	2.70%	3.22%	4.53%
C_1	4.09%	-	4.79	-
C_p	6.48%	6.15%	10.12%	13.30%
S	0.94	0.90	1.48	1.70

[1] Quantities are pounds per minute unless indicated otherwise.

CHAPTER VII

RESULTS AND DISCUSSION

A. Experimental Data and Calculations

Before the evaporator transient responses from the experiments outlined in Chapter VI are discussed, a brief look at the actual experimental data and the information it yields about the evaporator performance is of value. Tables number 6, 9 and 10 summarize the more important calculations from Appendix F. The flow rates and concentrations used in these calculations were, in every case taken from the original experimental data.

1. Steam Requirements

The steam flows at the steady state conditions before and at the end of each experimental run (Table 6) were calculated exactly as they were computed by the feedforward controller. That is, total steam requirements were considered to consist of steam to supply latent heat to the feed, to offset heat losses and to concentrate the feed to the product concentration. Thus the computed and measured steam flows can be compared to indicate the accuracy of the model. When this was done it was noted that the computed steam flow was consistantly higher than the actual steam flow. The average error was +5.48% and the actual error ranged from a low +0.22% to a high +11.1%. The reasons for this positive error in the model is closely associated with the actual computational errors inherent to the equipment into which the model was programmed. In Appendix D an error analysis indicated that the controller output was 6.70% below its theoretical output. Thus, during the initial

Table 6

SUMMARY OF HEAT TRANSFER AND CONCENTRATION DATA FOR EXPERIMENTS

Run No.	Steam Rate - lbs./min.				Percent Error in Steam Flow		Heat Transfer Coefficient Btu/hr.ft.^{2}°F			
	Measured		Computed				First Effect U_1		Second Effect U_2	
	Start	Finish	Start	Finish	Start	Finish	Start	Finish	Start	Finish
10L	1.322	1.322	1.403	1.369	5.76	3.44	255	199	450	410
20L	1.360	1.563	1.530	1.672	11.10	6.51	274	286	433	465
30L	1.410	1.410	1.496	1.515	5.75	6.31	279	288	440	441
40L	1.400	1.400	1.525	1.493	8.20	6.22	284	279	430	424
5FF	1.194	0.840	-	-	-	-	205	177	502	402
6FF	1.035	0.921	1.105	1.001	6.34	8.00	208	192	445	401
7FF	0.915	1.316	1.006	1.388	9.05	5.20	201	240	405	479
8FF	1.315	1.149	1.386	1.225	5.03	6.20	246	216	492	490
9FF	1.149	1.470	1.218	1.554	5.67	5.40	210	258	487	414
10FF	1.388	1.292	1.469	1.366	5.52	5.41	269	236	475	490
11FFB	1.260	1.230	1.332	1.265	5.40	4.35	254	224	522	505
12FFB	1.194	1.118	1.281	1.195	6.74	6.45	211	212	455	432
13FFB	Unstable									
14FFB	Unstable									
15FFB	1.001	1.360	1.093	1.455	8.40	6.53	224	263	413	477
16FFB	1.242	1.450	1.250	1.500	.53	3.33	134	242	470	474
17FB	1.050	1.357	1.079	1.367	2.69	.73	171	208	429	491
18FB	1.355	1.610	1.358	1.681	.32	4.21	208	241	508	490
Averages:					5.48		240		461	

experimental runs the controller output was tuned to overcome this deficit in its output. The result was that the K_s setting was too high, and theoretical steam consumption calculations based on this term were slightly excessive. Experimental errors in temperature and flow readings would then cause the error to vary to either side of the circuit error, which appears to have been the case.

The ratio of fresh steam flow to water evaporated (K_s) was calculated, by the procedure in Chapter II, for average experimental conditions to be 0.505. K_s was later balanced at 0.500 and left at this setting. However, possible subcooling of the condensate streams below the average vapor temperatures used in the calculation of K_s (220° in the first effect, 190°F in the second) would release more heat per pound of steam supplied, and could result in a value for K_s somewhat below 0.500. These observations support the explanation that the tuned value of K_s was slightly high and the cause of the descrepancy between calculated and measured steam flow rates.

A potential source of error in the actual steam rate resulted from the error in the programmed steam flow calibration curve. As Figure B-3 indicates, this was essentially a bias error and was therefore compensated for during the balancing of the feedforward controller prior to each run. This error, if left uncorrected, would have been quite large. In Table 7 the actual steam bias, the heat losses, and the excess bias are presented as equivalent steam flows for the runs in which the feedforward controller was used. The excess bias is the correction which was made to the steam flow to compensate for the error in its programmed calibration curve. It was essentially constant indicating

Table 7

STEAM BIAS SETTINGS FOR EXPERIMENTAL RUNS INVOLVING THE FEEDFORWARD CONCENTRATION CONTROLLER

Run No.	Actual Bias lbs./min.	Heat Losses lbs./min.	Excess Bias lbs./min.
6FF	0.164	0.082	0.082
7FF	.212	.086	.126
8FF	.166	.072	.094
9FF	.166	.067	.099
10FF	.166	.078	.088
11FFB	.166	.078	.088
12FFB	.127	.058	.069
15FFB	.132	.040	.092
16FFB	.132	.023	.109

that a bias adjustment was sufficient to overcome the calibration error. The controller model attempted to represent many process characteristics by three parameters (K_s, B, and L). Since some of these characteristics, such as steam pressures and solution temperatures, vary with changing operating conditions, it was not possible for the model to describe the evaporating system exactly. When this fact and the possibility of computational errors within the model are considered, it is not unreasonable to expect disagreement between theoretical and measured steam flows of the order of magnitude that is reported in Table 6.

2. Heat Transfer Coefficients

The overall heat transfer coefficients were calculated for each effect so that they might be quantitatively compared to those reported at each effect by Andre [(2), Appendix 8]. It was hoped that this comparison might indicate the effect which the process changes

described in Chapter IV had on the evaporator performance. The heat transfer areas used for the calculation of the coefficients were defined as the total wetted surface areas contacting the sugar solutions. In the first effect this included internal tube area, downcomer area and steam chest top and bottom plate areas (9.88 ft.^{2}), while at the second effect, this figure was the internal area of the three heating tubes (4.10 ft.^{2}). Andre's figures for the first and second effect heat transfer areas were 9.30 and 3.90 ft.^{2} respectively.

The heat transfer coefficients from each experimental run are listed in Table 6: the average coefficients are 240 and 461 Btu/hr.ft.^{2}°F for the first and second effects respectively. The average of Andre's figures (based on his definitions of heat transfer areas) is 379 and 358 Btu/hr.ft.^{2}°F for the first and second effects. The increase in the second effect coefficient can confidently be attributed to the increased fluid velocity in the heating tubes, which resulted following the installation of the new recirculation line and a high capacity circulation pump. The increased circulation velocity was able to improve the heat transfer at the second effect by lowering the heat transfer resistance within the heating tubes of the FCE. The temperature driving force required at each effect to maintain the flow of heat through the unit gives a direct indication of the heat transfer resistance inherent to each effect. The average heat transfer coefficients and temperature differences for each effect, are presented in Table 8 from Andre's study and from the present work. Since the heat load was approximately constant for all runs, it is evident that the resistance to heat transfer in the second effect, has been decreased to the point

where it is now slightly below the heat transfer resistance in the first effect. Since the average temperature difference at the second effect has been decreased, all input disturbances affecting the first effect boiling temperature (and therefore the second effect temperature difference) will have a greater effect on the transient concentration response in the second effect. The comparison of previous and current experimental transient data later in this chapter confirms this observation.

Table 8

AVERAGE EVAPORATOR HEAT TRANSFER CHARACTERISTICS

	Temperature Differences °F			Heat Transfer Coefficients Btu/hr.ft^2°F	
	First Effect	Second Effect	Overall	First Effect	Second Effect
Andre	23	42	65	379	358
Present Study	33	31	64	240	461

Attempts to correlate the heat transfer coefficients in each effect with the steam flow rates or steam chest pressures were unsuccessful. Other parameters which could have had an effect on the heat transfer coefficients are liquid concentration and level in the heating tubes (21) and the amount of inerts in the steam chests. The latter parameter may have changed from run to run following slight process modifications. For example, a leaking line on the first effect level guage was replaced following runs 20L, 30L and 40L (steam trap was also installed at this time). Just before run 16FFB another air leak was

discovered and sealed at the first effect. Following both these changes the first effect heat transfer coefficient fell and the second effect coefficient increased. This was reasonable since a decrease in the inerts content of the second effect steam chest would result in the shifting of a portion of the overall temperature difference from the second to the first effect. The increased operating temperature difference at the first effect and the fact that the heat transfer area has been defined as 9.88 ft.^{2} as opposed to 9.3 ft.^{2} in the previous study, explain the observed decrease in the first effect heat transfer coefficient during this study. It generally appears that the heat transfer coefficients in both effects increased as the flow of steam through the evaporator was increased. The unknown inter-dependence of these coefficients on other process variables makes it impossible to obtain a direct steam flow-heat transfer coefficient correlation.

3. Heat and Material Balances

The material balance errors, expressed as a percentage of the input stream flow, and the heat balance errors and heat losses in Btu/min., are presented for each experimental run in Tables 9 and 10 respectively. The overall heat balance provides a check on the assumptions which were made in calculating the heat and material balances. This balance includes as outputs the heat losses calculated in the balances around each effect. If the flows, temperatures and assumed temperatures were correct for each individual balance, there would be no error in the overall balance. Runs number 16FFB and 18FB were stopped before steady state conditions were achieved and it is just coincidental that the

Table 9

SUMMARY OF ERRORS IN MATERIAL BALANCE CALCULATIONS

Run No.	Mass Balance Errors %							
	First Effect		Second Effect		Overall		Sugar Balance	
	Start	Finish	Start	Finish	Start	Finish	Start	Finish
10L	-.15	-.15	-.10	1.31	-.21	.33	1.24	4.63
20L	.33	.98	-1.00	-4.80	-.36	-2.12	7.00	0.0
30L	1.55	1.54	-.52	0.0	1.21	1.54	-3.42	2.91
40L	1.20	1.88	1.03	1.65	2.56	3.14	9.43	3.46
5FF	1.01	.84	-1.23	0.0	.17	.84	6.66	4.42
6FF	.07	.54	4.31	.85	2.74	2.54	4.44	3.45
7FF	1.91	1.50	-.30	.25	1.67	1.64	3.25	4.25
8FF	2.02	1.21	1.06	2.38	2.72	2.88	6.70	8.05
9FF	.50	0.0	2.16	1.65	2.02	1.17	4.60	2.90
10FF	1.01	.67	-1.59	.26	0.0	.84	2.36	1.57
11FFB	1.01	1.68	2.90	1.22	2.62	2.52	9.30	1.45
12FFB	.84	.61	.39	.24	1.10	.79	5.30	.98
15FFB	-1.37	0.0	-6.94	1.42	-6.35	1.03	-11.20	4.85
16FFB[a]	1.35	1.01	0.0	1.19	1.35	1.75	-1.73	6.80
17FB	-.33	1.50	+.46	-.77	0.0	1.0	-1.33	-4.11
18FB[a]	-.33	-1.00	1.32	1.56	.50	0.0	-5.85	7.16

---

[a] Did not reach steady state.

Table 10
SUMMARY OF ERRORS IN HEAT BALANCE CALCULATIONS

Run No.	Heat Balance Error (Losses) Btu/min. Inter-Stage + First Effect Start	Heat Balance Error (Losses) Btu/min. Inter-Stage + First Effect Finish	Second Effect Start	Second Effect Finish	Overall Heat Balance Btu/min. Input Start	Input Finish	Heat Losses & Output Start	Heat Losses & Output Finish	Error Start	Error Finish
10L	47	17	54	21	1495	1491	1491	1491	-4	0
20L	130	165	98	45	1536	1732	1535	1722	-1	-10
30L	147	124	125	109	1577	1577	1581	1573	4	-4
40L	142	139	96	91	1571	1613	1577	1627	6	14
5FF	-35	-41	93	46	1363	1031	1390	1032	27	1
6FF	39	39	34	52	1214	1107	1233	1111	19	4
7FF	24	6	73	68	1102	1483	1108	1488	6	5
8FF	13	8	64	57	1494	1331	1496	1341	2	10
9FF	13	24	51	44	1334	1696	1332	1698	-2	2
10FF	24	1	55	75	1561	1468	1562	1478	1	10
11FFB	28	13	53	61	1441	1394	1451	1404	10	10
12FFB	-4	-13	69	62	1355	1302	1358	1303	3	1
15FFB	-22	-9	66	43	1197	1606	1168	1607	-29	1
16FFB[a]	-23	33	-4	-16	1401	1595	1405	1601	4	6
17FB	31	20	-4	26	1236	1527	1246	1529	10	2
18FB[a]	19	106	12	-18	1527	1799	1524	1795	-3	-4

[a] Did not reach steady state.

errors in these balances were quite small at the final operating conditions. An adjustment to the product valve prior to run 15FFB resulted in a separator level and product flow upset which carried over into the run. The deficiency in product flow at the start of this run is easily seen in the material and heat balances. The reason for the excess heat output in the overall balance for run number 5FF and 6FF is not indicated by the other balances, which appear normal for those runs.

There are three probable sources of error in the overall heat balance. They are the product and feed flow rates and the temperature of the condensate leaving the second effect steam chest, which was assumed equal to the condensation temperature. The steam flow and the condensate flows are used to calculate the heat losses from the evaporator--any errors in these quantities therefore cancels out when heat losses are included in the overall heat balance. The material balances show that the feed flow calibration may have been slightly low; the product flow calibration slightly high. The latter is especially evident in the sugar balance, which was usually high. Since this balance involved the product of two quantities (flow rates multiplied by concentrations) which were both subject to error, it was expected that the errors in this balance would be somewhat greater than observed in other balances. Finally, it was quite likely that the second effect condensate temperature would normally lie somewhere between the condensing temperature and the solution temperature in the heating tubes. Each of these possibilities would have the same net effect on the heat balance. The heat input would be slightly low or the calculated output too high, and there would be a positive error in the overall heat balance. In an

extreme case, the normal condensate flow of 0.9 pounds per minute subcooled 30°F to the second effect solution temperature, would alone cause an error of +27 Btu/min. in the overall enthalpy balance. This maximum error is approximately 2% of the normal heat input to the system. Considering the fact that all the calculations were made using the actual experimental data, the size of the errors in the system mass and energy balances is not at all unreasonable. Even though early calculations indicated slight feed and product flow calibration errors, it was felt that the accuracy of the calibrations concerned could not be improved by making new calibration runs. The check calibrations which were made from time to time during the experimental program did not indicate any significant errors in the curves reported in Appendix B.

It is evident from Table 10 that there was a noticeable decrease in the first effect heat losses following run 40L. It was at this time that the steam trap was added in the steam supply line. It appears that water entrainment in the steam could have thrown the flow measurement into error, so that an excessive steam flow was indicated. This excess would have appeared as heat losses when the heat balance was made around the first effect. Therefore, in addition to insuring the flow of dry steam to the first effect, the steam trap also removed a possible source of error in the steam flow measurement.

B. Evaporator Transient Responses

1. Experiments Conducted

Table 11 outlines the disturbances which were used to obtain the open loop and controlled evaporator transients presented in this chapter. The experiments are listed according to the type of product

Table 11

EXPERIMENTS: DISTURBANCES TO FEED CONCENTRATION, FEED FLOW AND STEAM FLOW RATES

Control System	Run No.	Disturbance to C_o	Refer To	Run No.	Disturbance to W_o	Refer To	Run No.	Disturbance to S or C_{psp}	Refer To
Open Loop	30L	4.32% to 3.06%	Fig.14	40L	2.92 to 3.72	Fig.15	20L	1.370 to 1.563	Fig.17
				EX-1	2.95 to 3.49	Fig.15	EX-3	1.42 to 1.54	Fig.17
				EX-2	3.80 to 3.40	Fig.15			
Conventional Feedback Concentration Control	17FB	4.53% to 3.08%	Fig.14	18FB[a]	3.00 to 3.48	Fig.16	EX-6	7.57% to 8.74% (1.32 to 1.49)	Fig.18
				EX-4	2.43 to 2.93	Fig.16			
Feedforward Concentration Control Via Overall Material Balance	6FF	3.10% to 3.43%	Fig. G-4	9FF	2.98 to 3.85	Fig.16	7FF	6.10% to 10.00% (1.131 to 1.518)	Fig.G-7
	8FF[b]	3.45% to 4.25%	Fig.14				10FF	10.50% to 8.50% (1.580 to 1.504)	Fig.18
Feedforward & Feedback Concentration Control Combined	11FFB[b]	3.44% to 3.90%	Fig. G-12	13FFB	2.98 to 3.87	Fig. G-13			
				14FFB	3.84 to 3.00	Fig. G-14	16FFB[a]	8.35% to 12.08% (1.458 to 1.638)	Fig.18
	12FFB[b]	3.38% to 3.80%	Fig.14	15FFB	3.00 to 3.87	Fig.16			

[a] Insufficient time to reach steady state.

[b] Includes a small simultaneous set point change due to "balancing" error.

NOTE: Quantities in brackets are steam flow changes called for to achieve a concentration set point change. All units are pounds per minute, unless indicated otherwise.

concentration control that was used and the disturbance that initiated the run. Experimental runs EX-1 to EX-4 and EX-6, which were conducted by Andre, are included in this chapter for comparison to similar runs from the present study. The transient product concentration responses for many of the runs listed in Table 11 are discussed generally in Table 12 and are presented collectively in Figures 14 to 18 inclusive. The "transient time" referred to in Table 12 was defined as the time required for the product concentration to reach and remain within 0.05% sugar of its final steady state value.

It is evident from Figures 14 to 18 that some of the experimental runs did not commence with the actual product concentration exactly on the "set point." The reason for this anomaly lies in the balancing procedure for the feedforward controller as described in Appendix C. Prior to each run, the computer parameters B and L were checked and reset if necessary, so that the evaporator could be switched to feedforward control without altering the steady state steam flow. The balancing procedure involved taking a sample of the product from the recirculation line, quenching it to the correct calibration temperature for the bench refractometer, setting the product refractometer output to correspond to the correct concentration, and then setting the product concentration set point on potentiometer P-3 in the feedforward controller at the value determined by analysis. Cooling the sample with tap water and using a thermometer to indicate the solution temperature was clearly not as accurate as allowing the sample to reach equilibrium in a constant temperature bath before analysis. However, the necessity for an immediate

Table 12

OBSERVATIONS ON THE TRANSIENT BEHAVIOR OF A DOUBLE EFFECT EVAPORATOR UNDER OPEN LOOP, FEEDFORWARD, FEEDBACK, AND FEEDFORWARD-FEEDBACK CONCENTRATION CONTROL

Control System	Disturbance to Feed Concentration	Disturbance to Feed Flow	Disturbance to Product Concentration Set Point (or Steam)
Open Loop	Transient Time: 160 min. Overdamped Response. Approximate Sensitivity: $\frac{\Delta C_p}{\Delta C_o} = 2.4$	Transient Time: 150 min. Overdamped Response. Approximate Sensitivity: $\frac{\Delta C_p}{\Delta W_o} = \frac{4.2\%\ \text{Sugar}}{\text{lb. Feed}}$	Transient Time: 175 min. Overdamped Response. Approximate Sensivity: $\frac{\Delta C_p}{\Delta S} = \frac{18.7\%\ \text{Sugar}}{\text{lb. Steam}}$
Conventional Feedback Concentration Control T_i = 7 min. T_d = 5 min. PB = 130%	Transient Time: 155 min. No Offset at Start or Finish. Highly Oscillatory Response.	Steady State Not Reached After 200 min. 0.02% Sugar Balancing Error at Start of Run. Highly Oscillatory Response. Andre (EX-4) Reported Steady State after 2 hr.	Run Not Made. Andre (EX-6) Reported Steady State After 140 min. Oscillatory Response.
Feedforward Concentration Control Via Overall Material Balance	Transient Time: 120 min. Offset from Set Point: Start: 0.20% Sugar Finish: 0.0 Overdamped Response.	Transient Time: 140 min. Offset from Set Point: Start: 0.01% Sugar Finish: 0.40% Sugar Oscillatory Response.	Transient Time: 115 min. Offset from Set Point: Start: 0.08% Sugar Finish: 0.19% Sugar Overdamped Response.
Feedforward and Feedback Concentration Control Combined. Feedback Control Action unless Otherwise noted is: T = 30 min. PB = 400%	Transient Time: 104 min. 0.07% Sugar Balancing Error at Start of Run. Oscillatory Response with T_i = 10 min. PB = 100%	Transient Time: 120 min. Oscillatory Response.	Steady State not Reached after 4 hr. 0.08% Sugar Balancing Error at Start of Run. Very Slow Oscillatory Response.

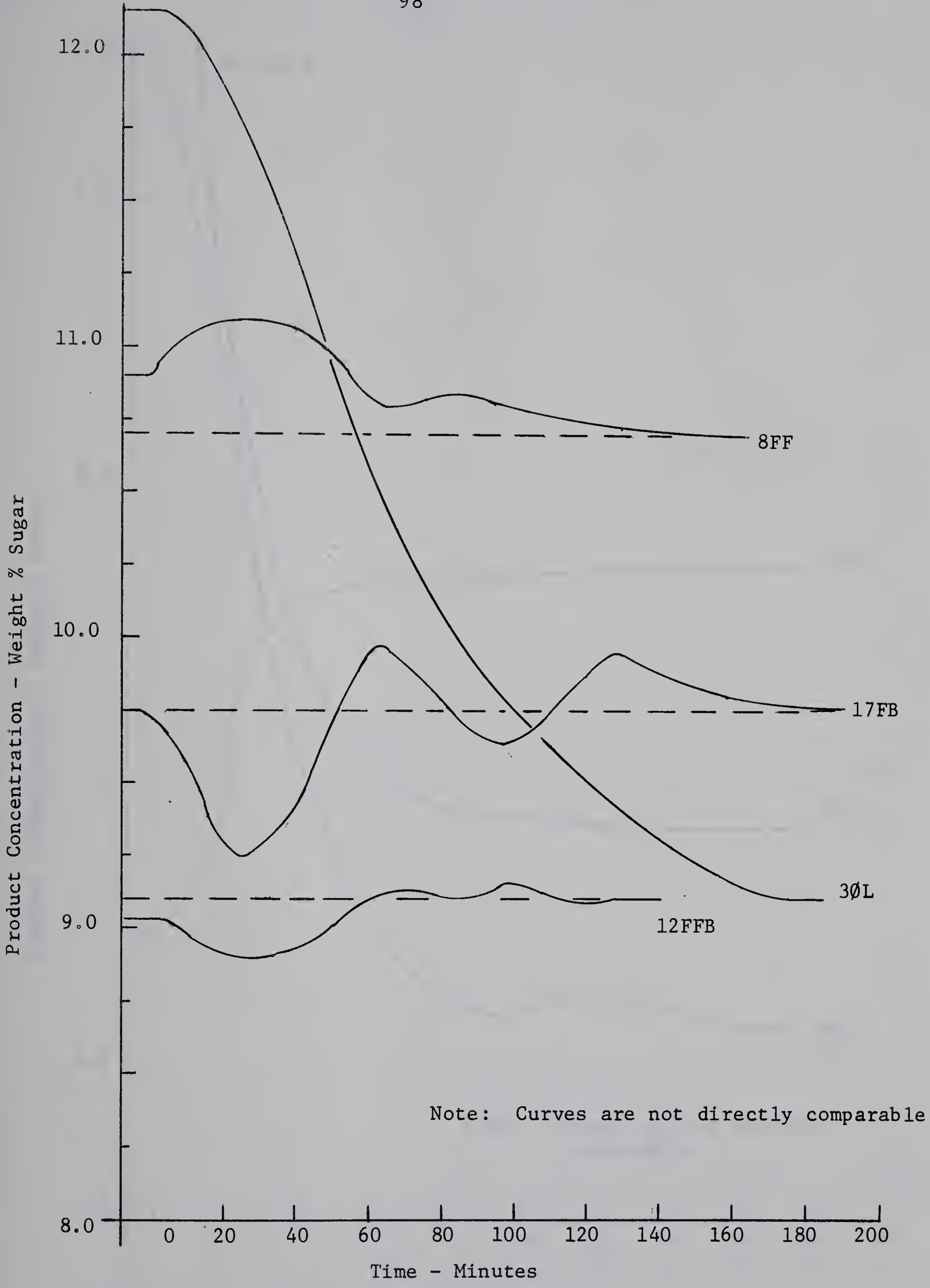


Figure 14 THE RESPONSE OF PRODUCT CONCENTRATION TO DISTURBANCES IN FEED CONCENTRATION

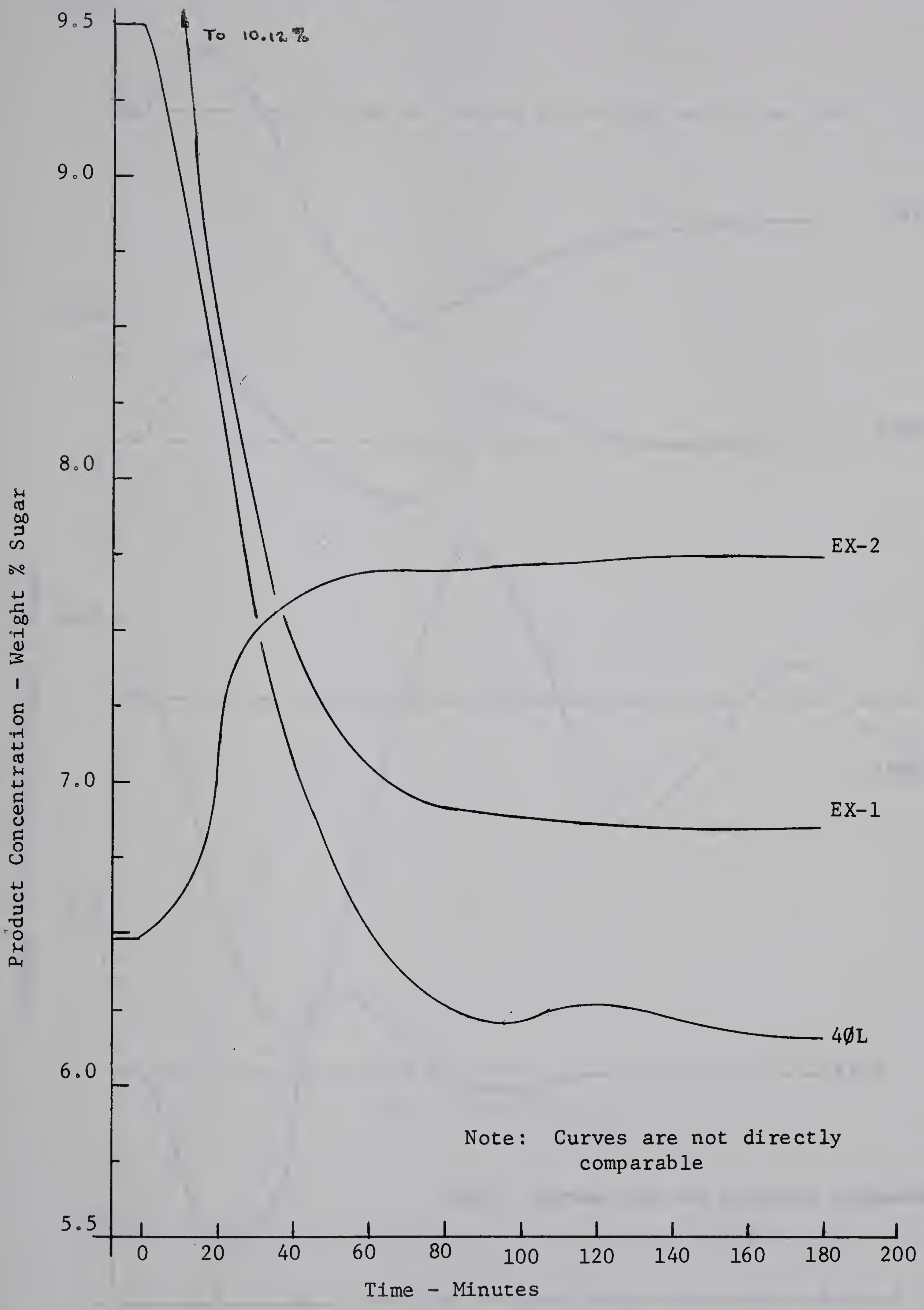


Figure 15 THE OPEN LOOP RESPONSE OF PRODUCT CONCENTRATION TO DISTURBANCES IN FEED FLOW RATE

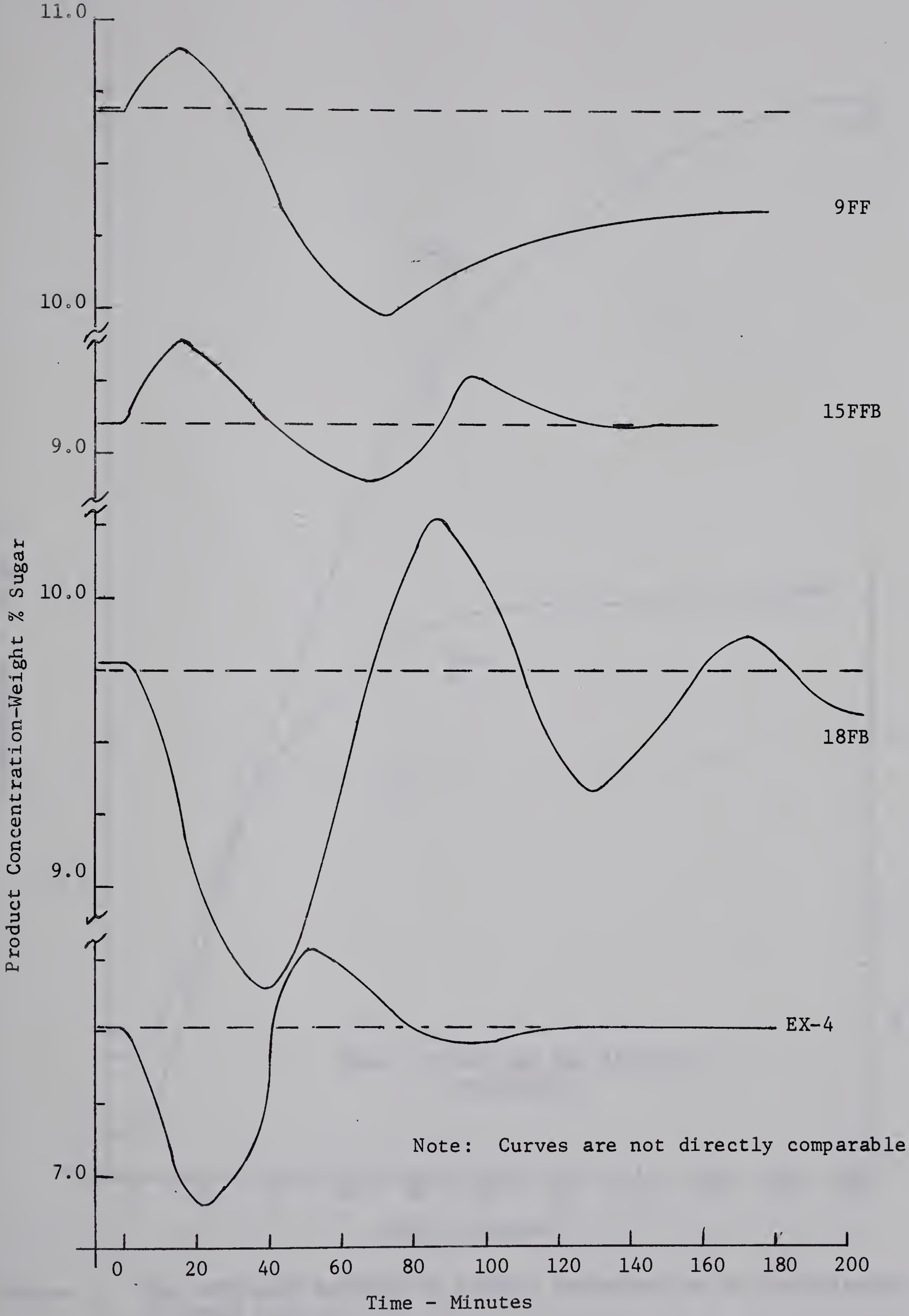
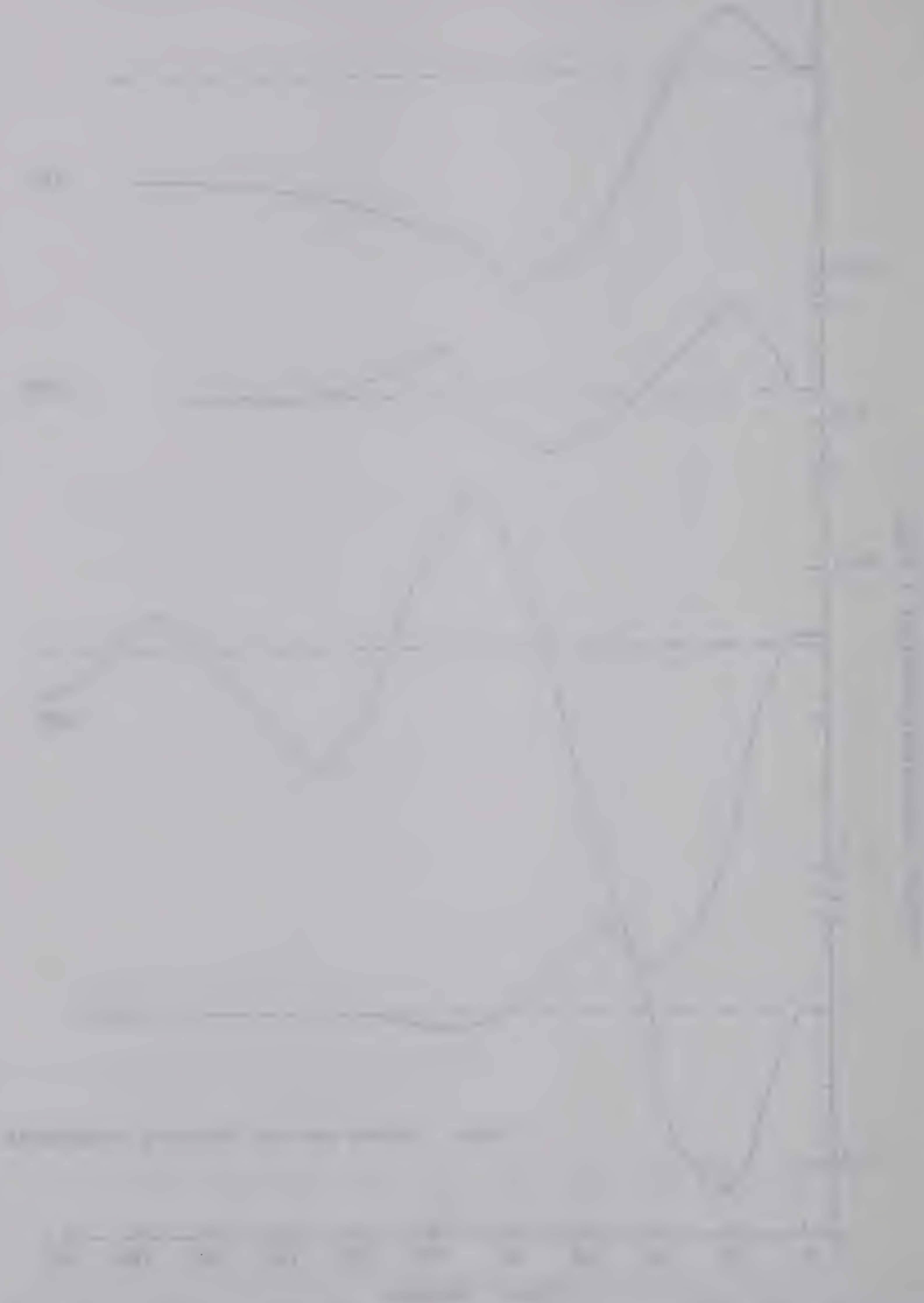


Figure 16 THE CONTROLLED RESPONSE OF PRODUCT CONCENTRATION TO DISTURBANCES IN FEED FLOW RATE



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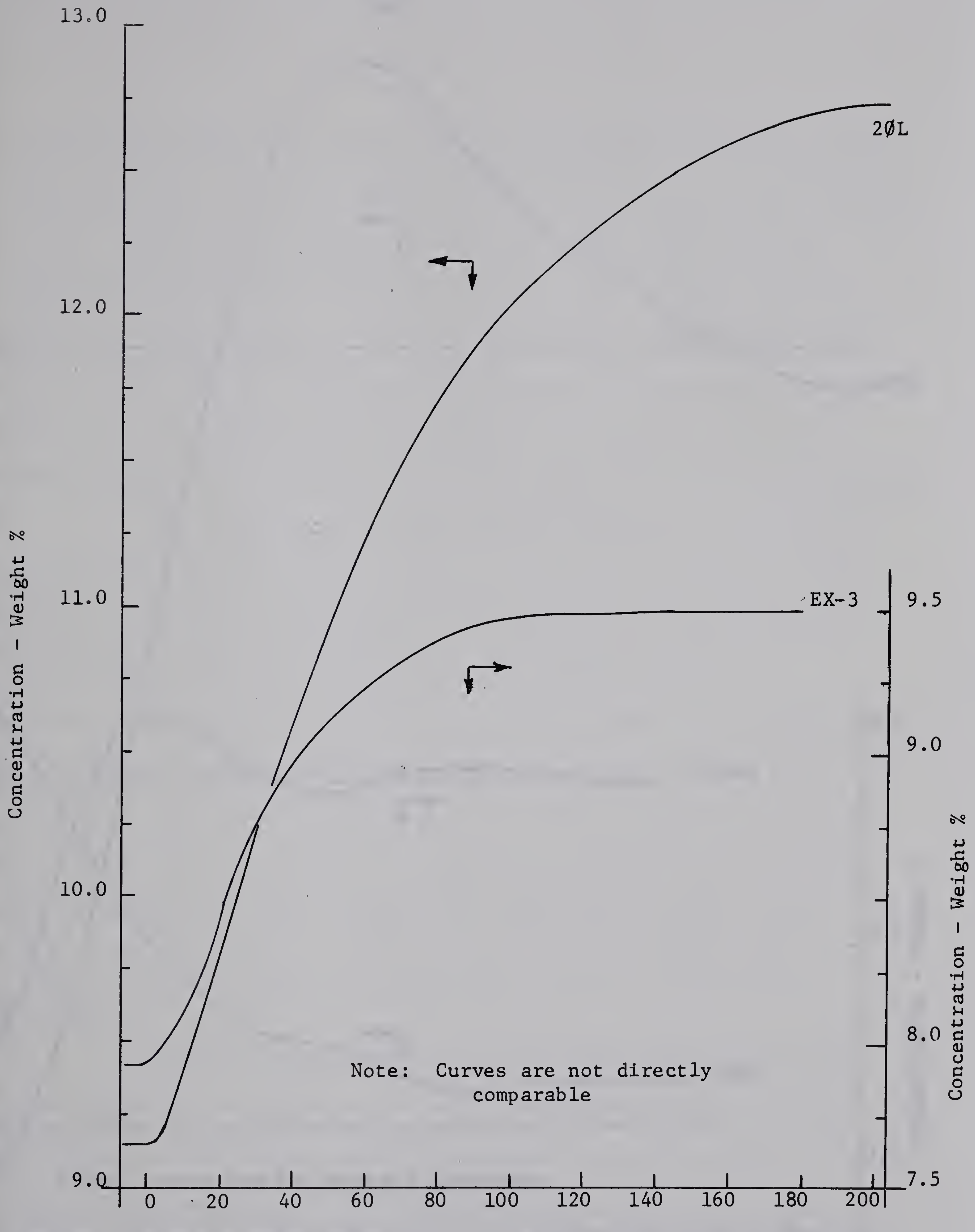
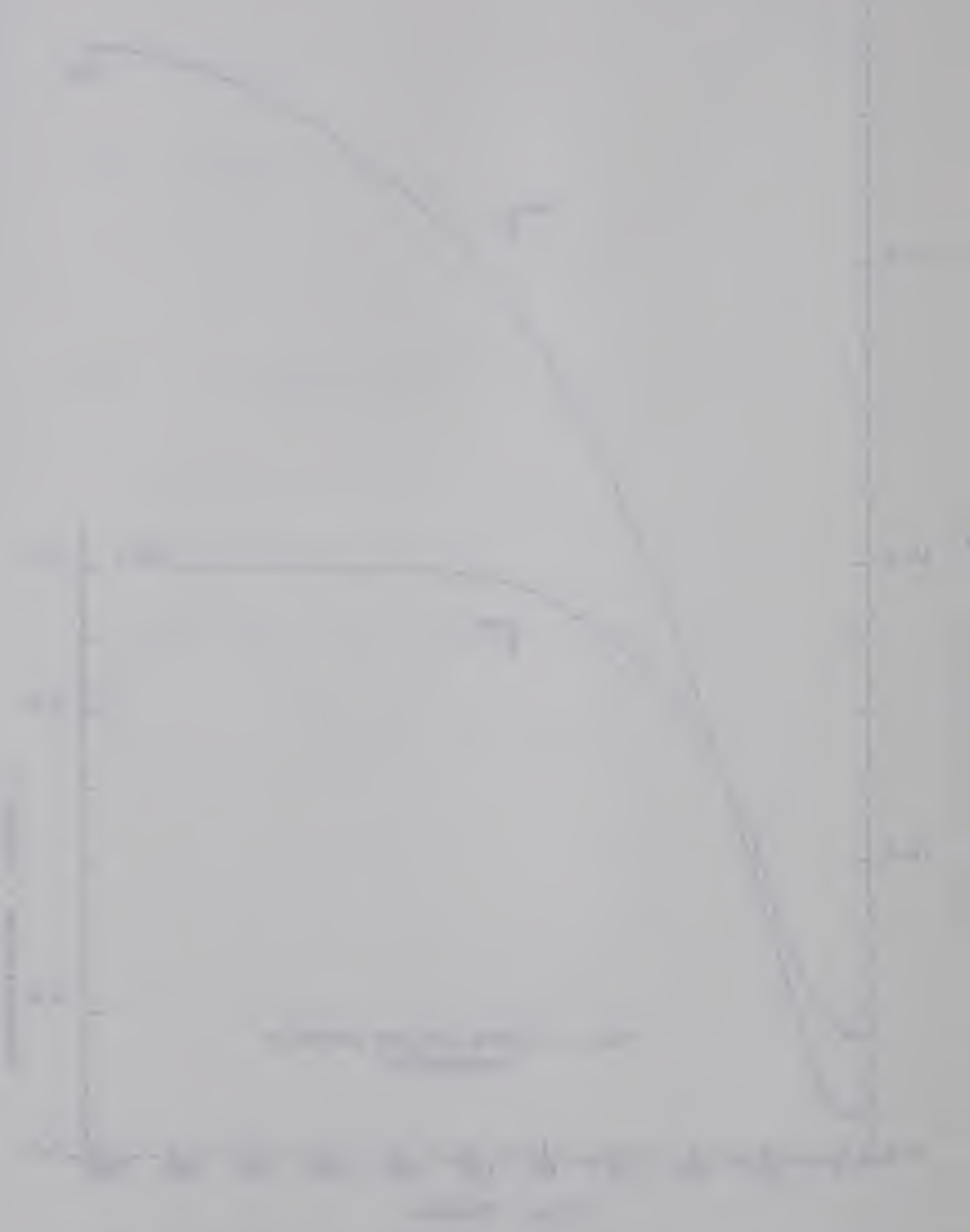


Figure 17 THE OPEN LOOP RESPONSE OF PRODUCT CONCENTRATION TO DISTURBANCES IN STEAM FLOW RATE



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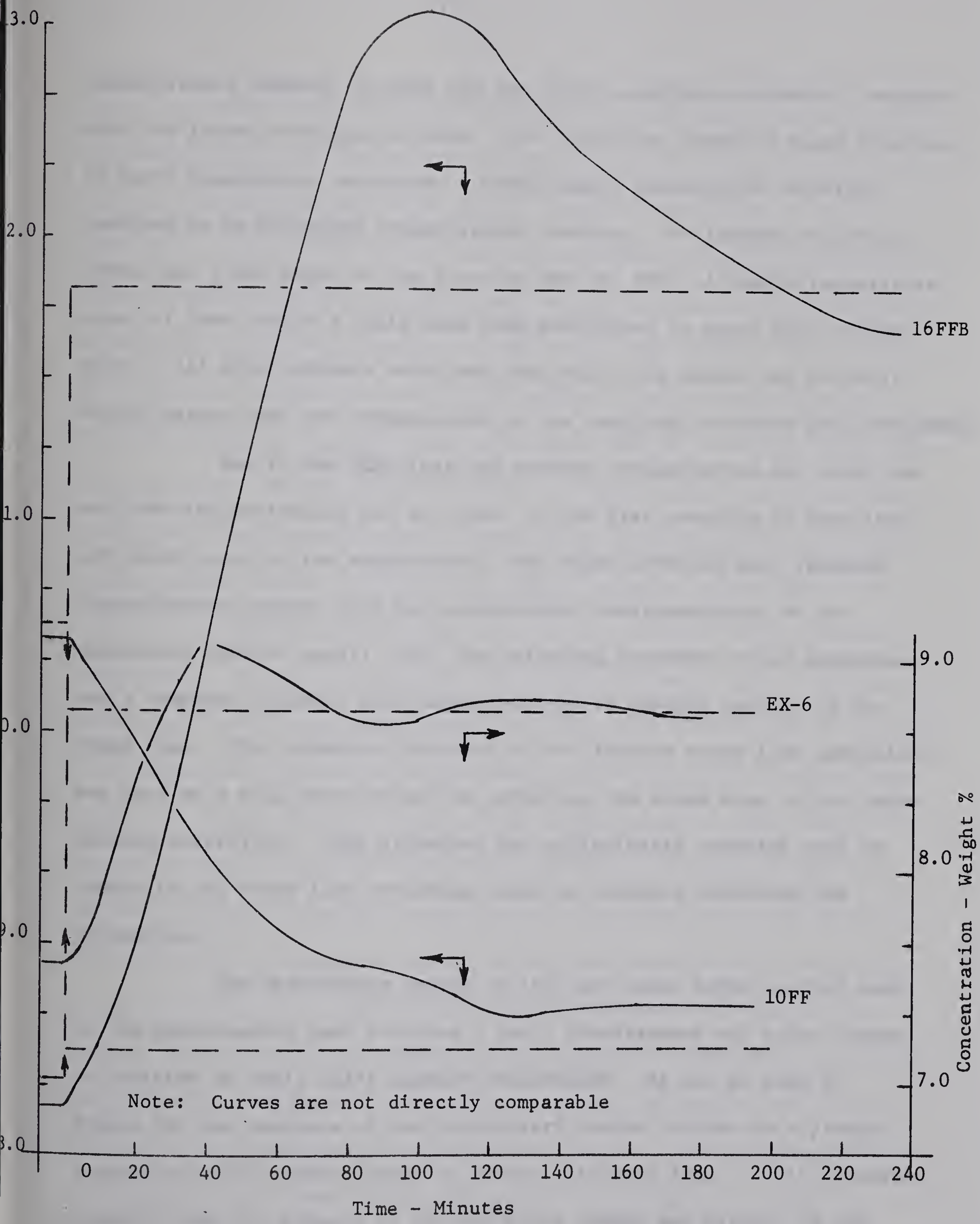


Figure 18 THE CONTROLLED RESPONSE OF PRODUCT CONCENTRATION TO SET POINT CHANGES

concentration reading, so that the set point could be programmed, required that the former technique be used. The refractive index of sugar solutions is quite temperature sensitive: a wrong sample temperature therefore resulted in an erroneous concentration reading. The largest set point offset was 0.20% sugar at the start of Run No. 8FF. A sample temperature error of less than 6°F would have been sufficient to cause this setting error. All other offsets were less than half this amount and are well within reason when the consequences of the sampling procedure are considered.

Due to the fact that the product concentration set point was set from the controller for all runs, it was also possible to have this set point error in the experimental runs which involved only feedback concentration control (via the conventional instrumentation on the evaporator control panel). All the balancing procedure could guarantee was a bumpless transfer from manual control to cascade control of the steam flow. The deviation indicator on the Foxboro steam flow controller was used as a null meter to aid in adjusting the steam bias in the feedforward controller. This procedure was sufficiently accurate that no change in the steam flow recording could be detected following the switch-over.

The unfortunate result of this set point error is that many of the experimental runs involved a small simultaneous set point change in addition to their major planned disturbance. As can be seen by Figure 18, the response of the feedforward control system to a product concentration set point change is essentially open loop. It is therefore possible that the effects of the set point change may persist in the transient response even longer than the effects of the major disturbance.

Any time a feedforward control system is used, there is a possibility that the controlled variable may not be returned exactly to the set point following a disturbance due to inherent errors in the controller or in the model used by the controller. Runs number 9FF and 10FF on Figures 16 and 18 respectively and runs number 6FF and 7FF in Appendix G are evidence of this possibility. Runs number 6FF and 7FF were the first runs made using feedforward control only and the final concentration offset is attributed to tuning errors in the K_s and heat loss bias settings. During run 9FF there was a significant increase in feed latent heat requirements, and in addition, a positive error in the feed concentration voltage as it left the delay circuit. Both these factors combined to produce a deficiency in steam flow at the final steady state conditions. Since the evaporator is very sensitive to steam flow (as is indicated by the approximate sensitivity for run number 20L in Table 12) the shortage of steam flow in run number 9FF resulted in a fairly large offset in the final steady state product concentration. The factors generally responsible for concentration offset errors during the feedforward runs are discussed in Appendix D. An attempt was also made to calculate the concentration offset error, which could result due to changing heat loss or latent heat steam requirements during the runs involving the use of the feedforward controller. Experimental errors in the data required for these calculations made an accurate estimate of the product concentration error impossible to obtain. Nevertheless, these calculations are summarized in Table D-2 to illustrate the possible

effect that even small errors in the feedforward controller settings may have on the actual product concentration.

2. Discussion of Experiments

Because of the set point offset which resulted at the beginning of some experimental runs, it is not possible to compare the responses of the various control systems in any other way than a qualitative manner. Criteria such as the decay ratio of successive peaks in the response curve, or the area within the transient curve cannot easily be applied to runs which were initiated by two simultaneous disturbances.

It is possible however, to detect some obvious trends in the transient response curves and in Table 12.

(a) Open loop runs

(i) The product concentration sensitivity is lowest for feed concentration disturbances, which are damped by the ten minute, first order time constants in each effect. The concentration is somewhat more sensitive to feed flow disturbances, although this sensitivity must depend upon the level control exercised on the holdups in the two effects. The evaporator heat transfer dynamics are quite rapid and make the product concentration highly sensitive to steam flow changes.

(ii) The evaporator response is slow and non-oscillatory having the appearance on an over-damped second order response. Approximately three hours is required to reach steady state conditions following a disturbance to C_o, W_o or C_{psp}.

(iii) The open loop responses obtained during the present study are not significantly different from those reported by Andre.

(b) Conventional feedback product concentration control runs.

(i) Feedback control runs reported by Andre reached steady state in approximately two hours. Using the same controller settings on the concentration and steam flow controllers as reported by Andre and initiating the runs with similar disturbances, resulted in stable but highly oscillatory product concentration responses. Since the product concentration response did not stabilize within four hours after a feed flow upset, a run involving the more sensitive product concentration set point change was not attempted.

(ii) If the amplitudes of the first overshoot in runs 18FB and EX-4 are divided by the magnitudes of the flow disturbances which caused the upsets, the "normalized" overshoot of run 18FB is nearly twice that of run EX-4.

Both (i) and (ii) above support the observation made earlier in this chapter regarding the decrease in the heat transfer resistance at the second effect, which followed the installation of a new circulation system. Since the operating temperature difference at the second effect has been decreased during this study, it is reasonable that any upset causing a change in the temperature of the supply steam from the first effect would now have a greater effect on the product concentration than it did previously when the temperature difference was higher.

(c) Feedforward concentration control runs.

(i) The ability of this control system to measure disturbances in the feed flow rate or concentration and offset much of their effect on the product concentration effectively shortened the transient time period following these disturbances.

(ii) Following a product concentration set point decrease, the transient response of the evaporator system was approximately one hour shorter than the response time following an open loop steam flow increase. This could not have been the result of the feedforward control system since it effectively called for an immediate step decrease in the steam flow, which would have been followed by an open loop type concentration response. It therefore must be concluded that the evaporator system is quite non-linear, and that the product concentration response to steam flow changes is not independent of the magnitude of the change or of the direction in which that change was made.

(iii) A feedforward control system using a steady state model will overcompensate for a disturbance in the transient period immediately following the detection of the upset in an input varialbe. This statement is illustrated by the concentration response of run 9FF (Figure 16) following a step increase in the feed flow rate. Whereas conventional feedback control must allow a negative concentration error before compensation for this disturbance can begin, the feedforward control system initially caused the product concentration to overshoot its set point value. There should

therefore be one combination of feedforward and feedback control which will minimize the concentration deviation following a feed flow disturbance. However, if the system model used by the feedforward controller accurately describes the dynamic and steady state characteristics of the evaporator, the feedforward compensation will be perfect,and the concentration will be held on its set point following any disturbance without the addition of feedback compensation.

When using a steady state model for a feedforward controller, the concentration response will normally fall smoothly to its set point value from the peak of the initial overshoot. The reason for the permanent concentration offset evident in the concentration curve of run 9FF has already been discussed in this chapter and in Appendix G. Unfortunately, a sticking valve in the product line also allowed a temporary level buildup in the separator. This allowed windup in the level controller and permitted product withdrawal at too rapid a rate when the valve ceased sticking. Thus the product concentration fell too low during the transient and had to approach its steady state position from below instead of above. Most of the value of run number 9FF was lost because of the appreciable set point change that accompanied the flow rate disturbance and due to the sticking valve. The run did reach its final steady state quite rapidly in spite of these interferences and so was not discarded. Since run number

15FFB showed the success of the feedforward controller in compensating for feed flow disturbances, it was not considered essential that run number 9FF be repeated.

The product concentration response following a feed concentration disturbance does not contain the over-correction indicative of a steady state feedforward controller. The reason for this lies in the addition of a ten minute time constant element and a one minute transportation lag (manually set) to the feed concentration measurement signal. These additions permitted the model to simulate the first effect concentration response to feed concentration upsets, and to adjust the steam flow accordingly when this concentration left its steady state value. Run number 8FF indicates that the feedforward controller initially undercompensated for a feed concentration disturbance: earlier runs including a five minute first order lag and no transportation lag resulted in immediate overcompensation. It appears that further tuning of the controller, involving trial-and-error adjustments to the time constant and dead time setting, might produce a concentration response which does not depart from the set point value following a feed concentration upset. Such a tuning program would have required extensive experimental time and was therefore not attempted.

(iv) Although an absolute comparison of the maximum product concentration deflections for the experimental runs

has not been made, it is evident that the feedforward control system was able to substantially reduce the amplitude of the primary oscillation, following a feed disturbance, from that encountered using feedback control alone. Because the feedforward controlled system response to set point changes is essentially the same as an open loop steam change response, the product concentration responds slowly and approaches the concentration set point asymptotically. This type of response is not usually desirable, and indicates that the model needs some form of dynamic compensating control in addition to the steady state steam adjustment if the product concentration response to set point changes is to be improved.

(d) Feedforward-feedback product concentration control runs.

(i) The addition of negative feedback to the feedforward control action insured that no product concentration offset could result at the final steady state conditions. As the last row in Table 12 indicates, the transient time period following disturbances in feed flow or composition was slightly shorter than the transient period when feedforward control only was used, and was considerably shorter than the corresponding periods when open loop or feedback control runs were made.

(ii) The feedback control action used was held to a minimum of 400% proportional band and thirty minutes per repeat for the majority of the feedforward-feedback controlled runs. This constitutes a very small amount of feedback trimming for

the feedforward controller output. It would therefore not be expected to cause a noticeable change in the feedforward concentration transient responses which closely follow the actual feed upset. A comparison of the first thirty minutes of transient response from runs 8FF and 12FFB or 9FF and 15FFB confirms this surmise.

(iii) It is evident from comparisons among runs 12FFB, 8FF and 17FB and among runs 15FFB, 9FF and 18FB (all involving feed disturbances), that the addition of feedback trimming to the feedforward control signal produced better product concentration control than either of the feedforward or feedback control systems could provide independently. The combined control system reduced both the magnitude of the deviations and the transient period following feed upsets, and prevented any permanent offset in the product concentration.

(iv) It was noted above that the feedforward controlled concentration response following a set point change was not very satisfactory: the addition of the small amount of feedback trimming noted above only made the response worse, as the concentration transient for run 16FFB indicates. Unfortunately this run involved a very big set point change of approximately 50%, but the response did indicate that the negative feedback added was not suitable. The reset windup, during the one hour period in which there was effectively a large product concentration error, caused a large overshoot. Because the feedback action is so small, many hours were

required before this overshoot was overcome and the product concentration began to approach the set point. This very slow response indicates that there should be no feedback integral action on the controller in the first hour of the transient period. This could be accomplished by locking out the integral action during this period and tuning it in as the concentration approached the set point.

3. Concentration, Flow, and Level Transient Responses

The transient responses of product concentration, first effect bottoms, steam, and product flows, and first and second effect liquid levels follow in Figures 19 to 51 inclusive for the majority of the experimental runs. The curves are largely self-explanatory and do not require individual discussion. Liquid levels were referred to the top of the steam chest and the separator bottom plate in the first and second effects respectively.

During early trial runs (before those reported in this chapter) the first effect level controller was set with the widest proportional band and least reset action required to maintain the liquid level within reasonable limits (85% P.B. and 20 minutes/repeat). The lower limit on the first effect level was minus four inches; the upper limit was 13 inches of liquid. These upper and lower limits were set to prevent the first effect from boiling dry, and to prevent liquid entrainment in the overhead vapor line and poor circulation respectively. Using the above settings the first effect level required approximately three hours to settle out following a disturbance, and the concentration transient

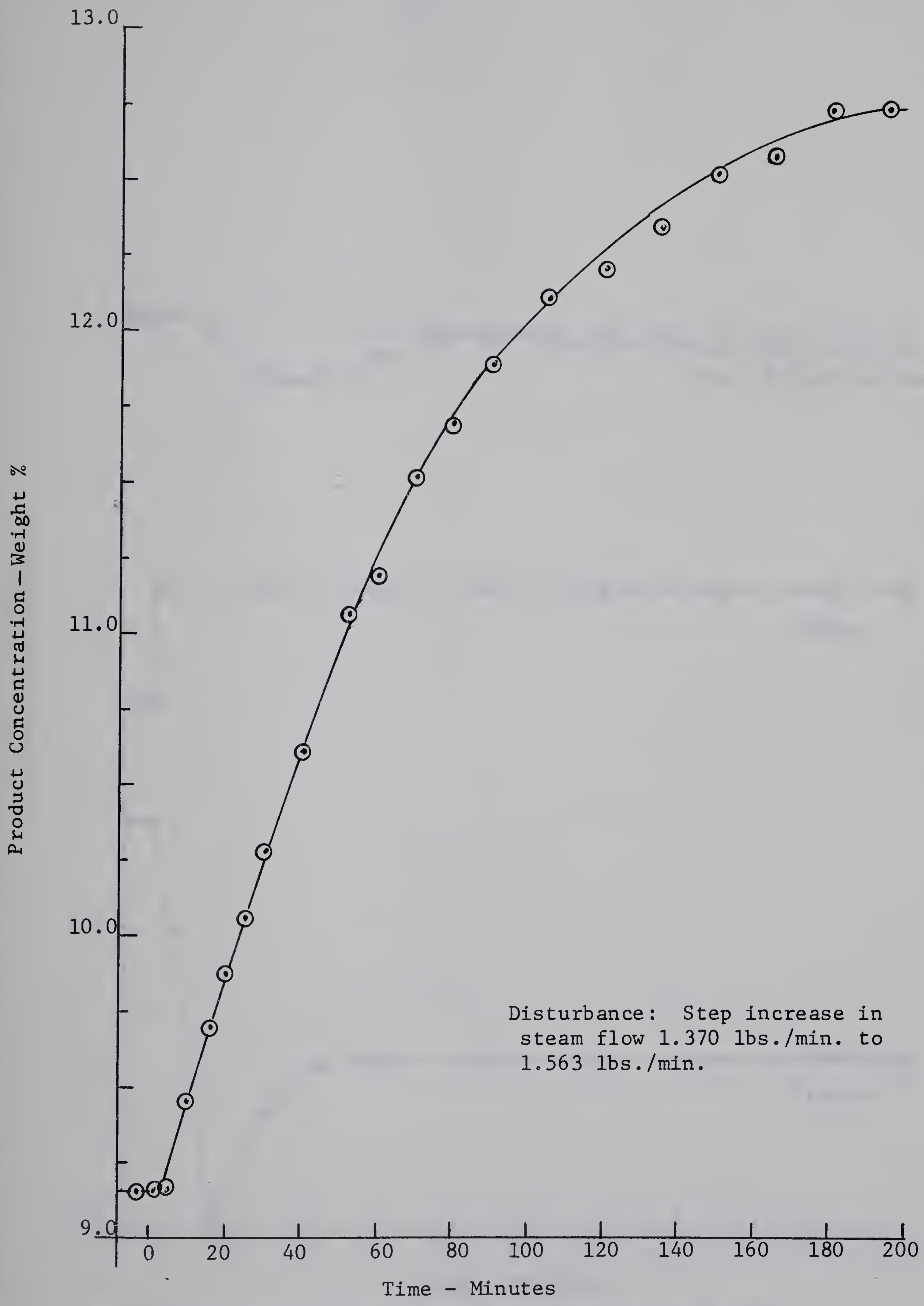


Figure 19 TRANSIENT CONCENTRATION RESPONSE FOR RUN NO. 20L

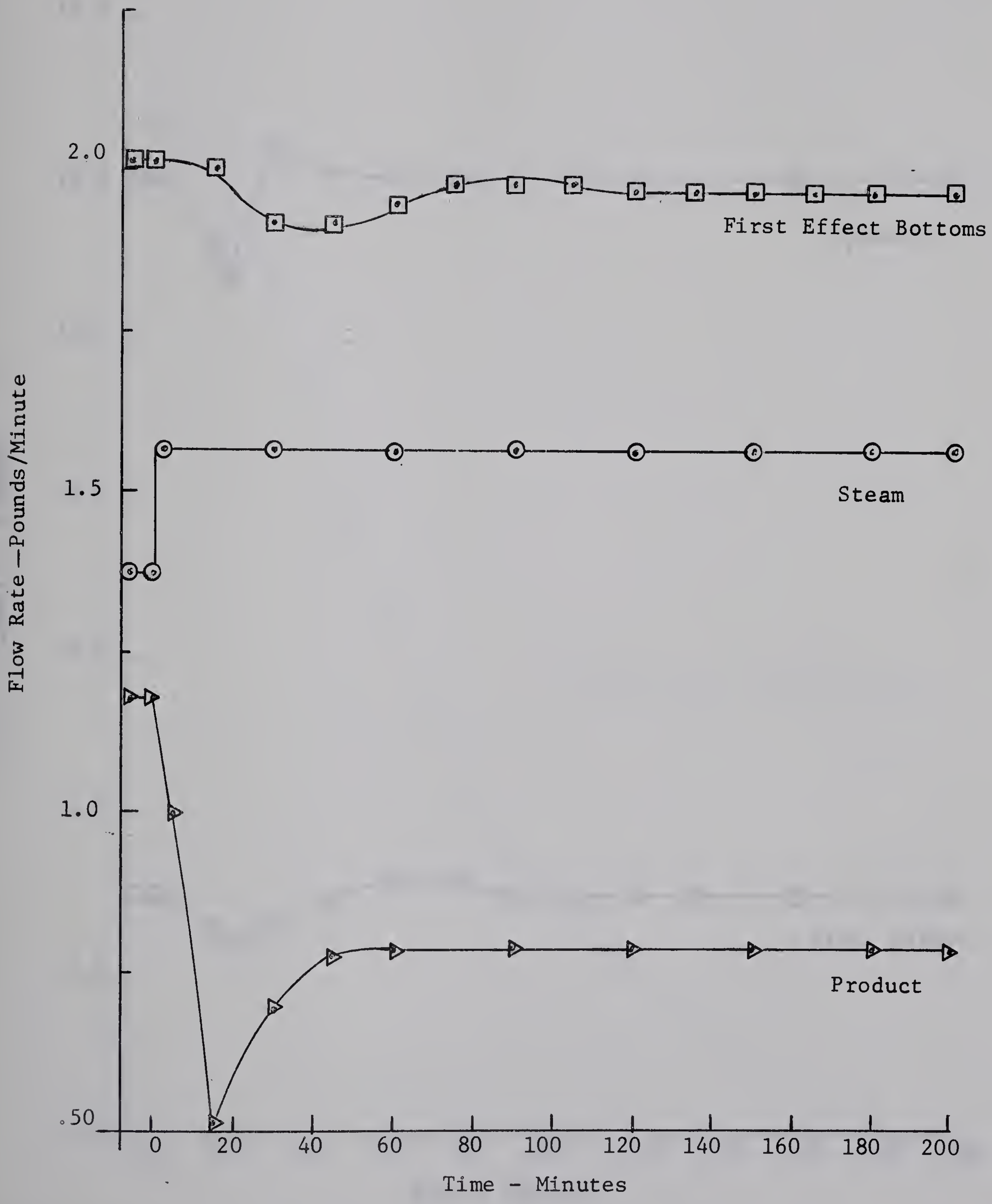
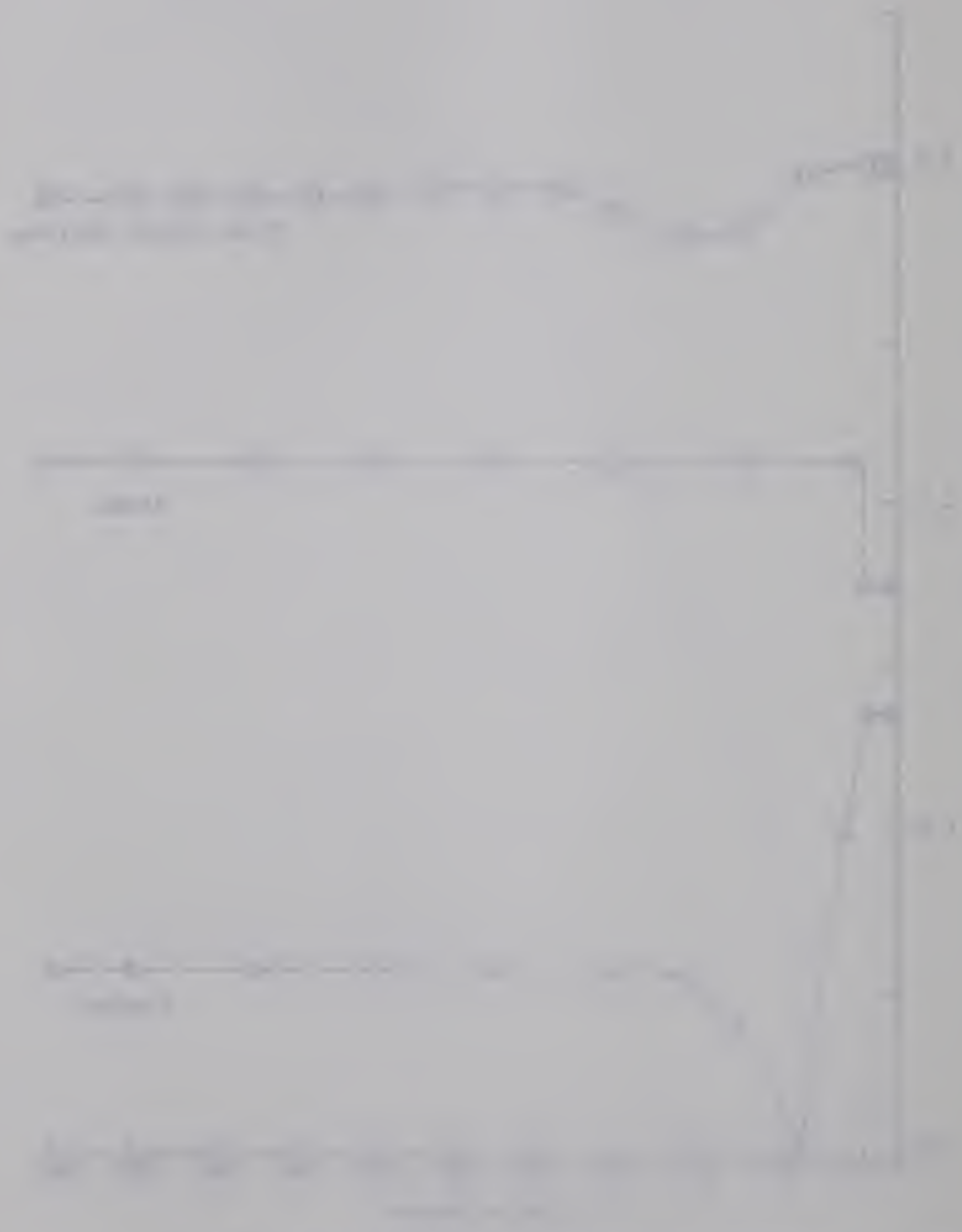


Figure 20 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN NO. 20L



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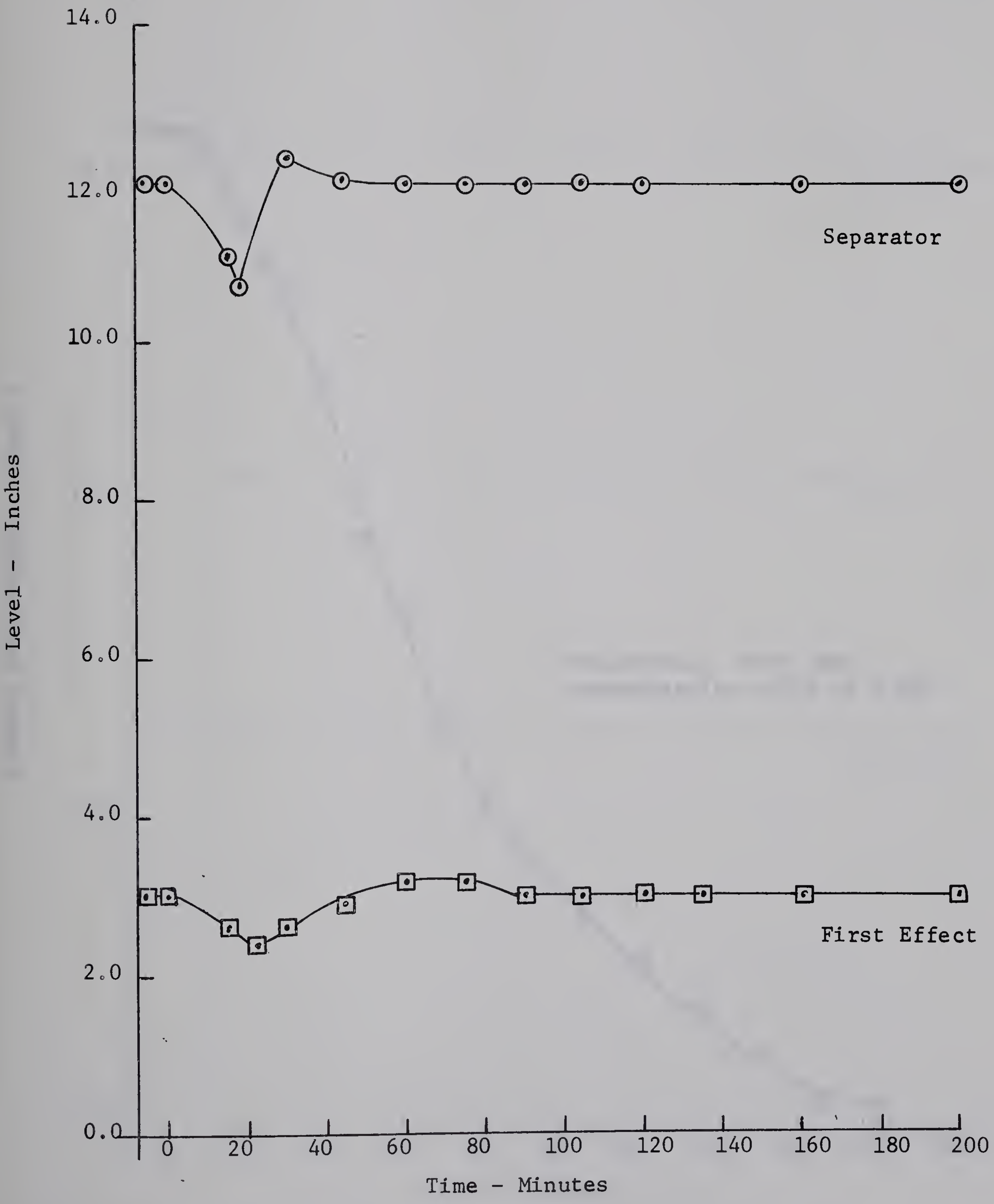

Figure 21 TRANSIENT RESPONSE OF SYSTEM LEVELS FOR RUN NO. 20L

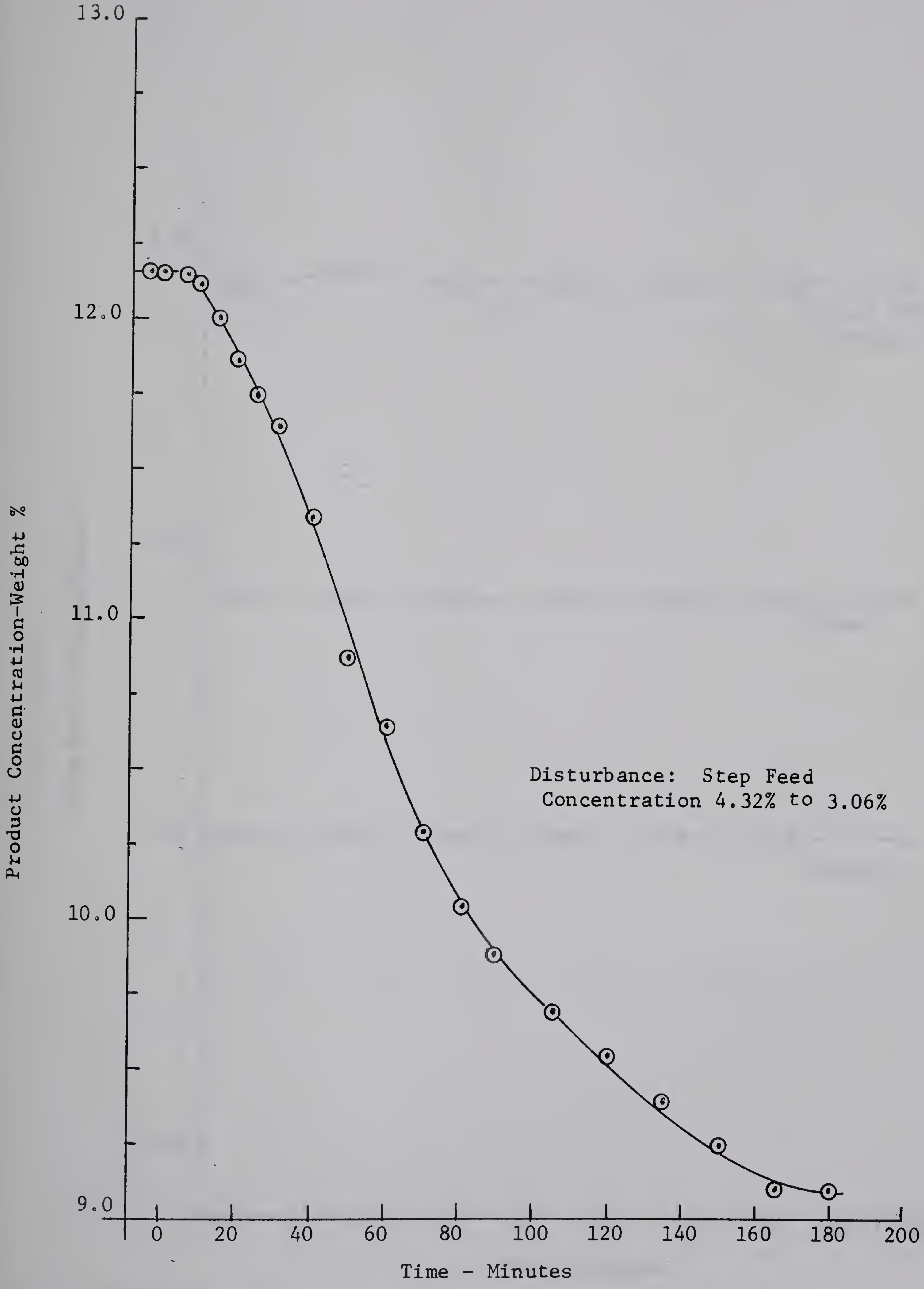

Figure 22 TRANSIENT CONCENTRATION RESPONSE FOR RUN NO. 30L

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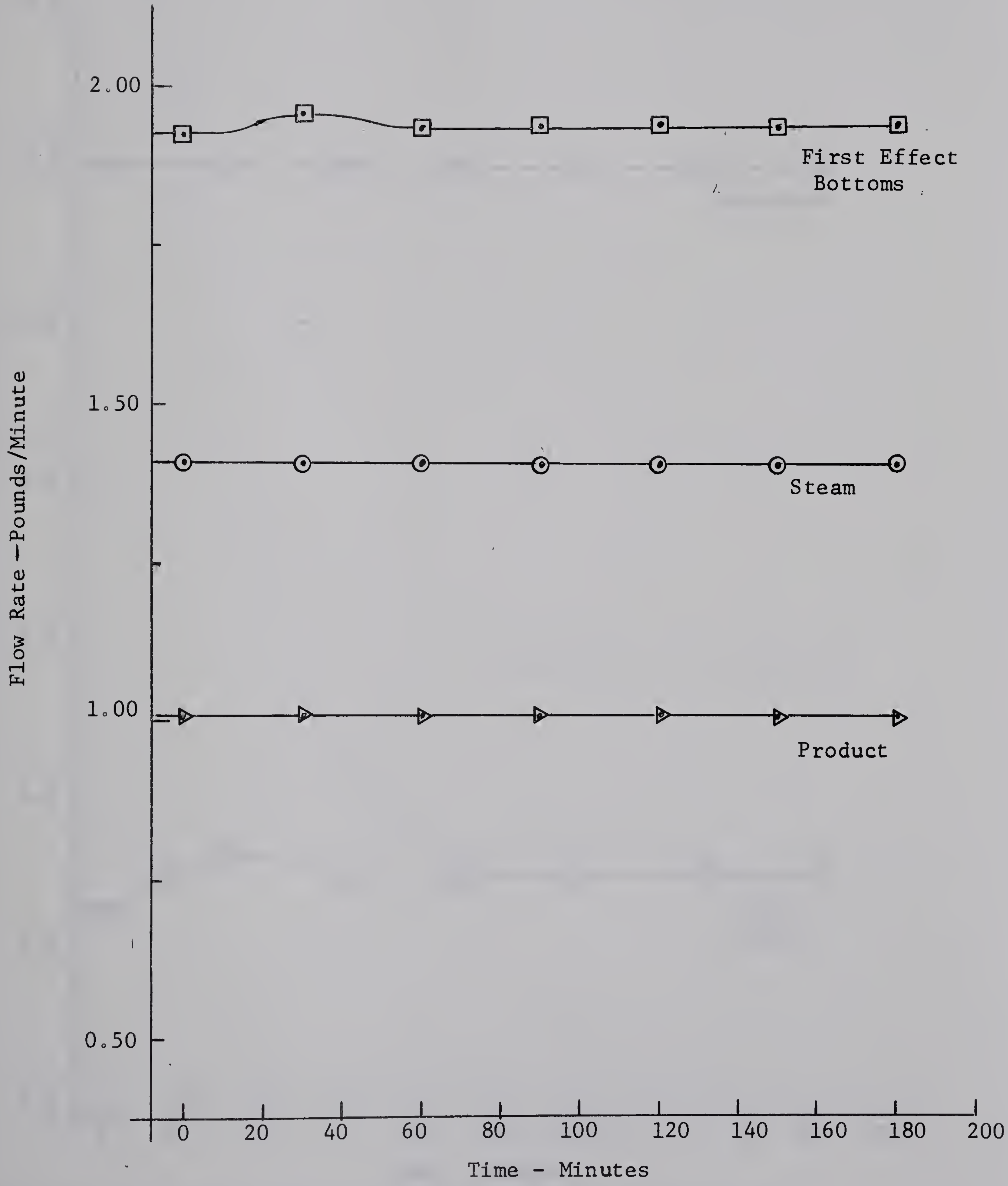


Figure 23 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN NO. 30L

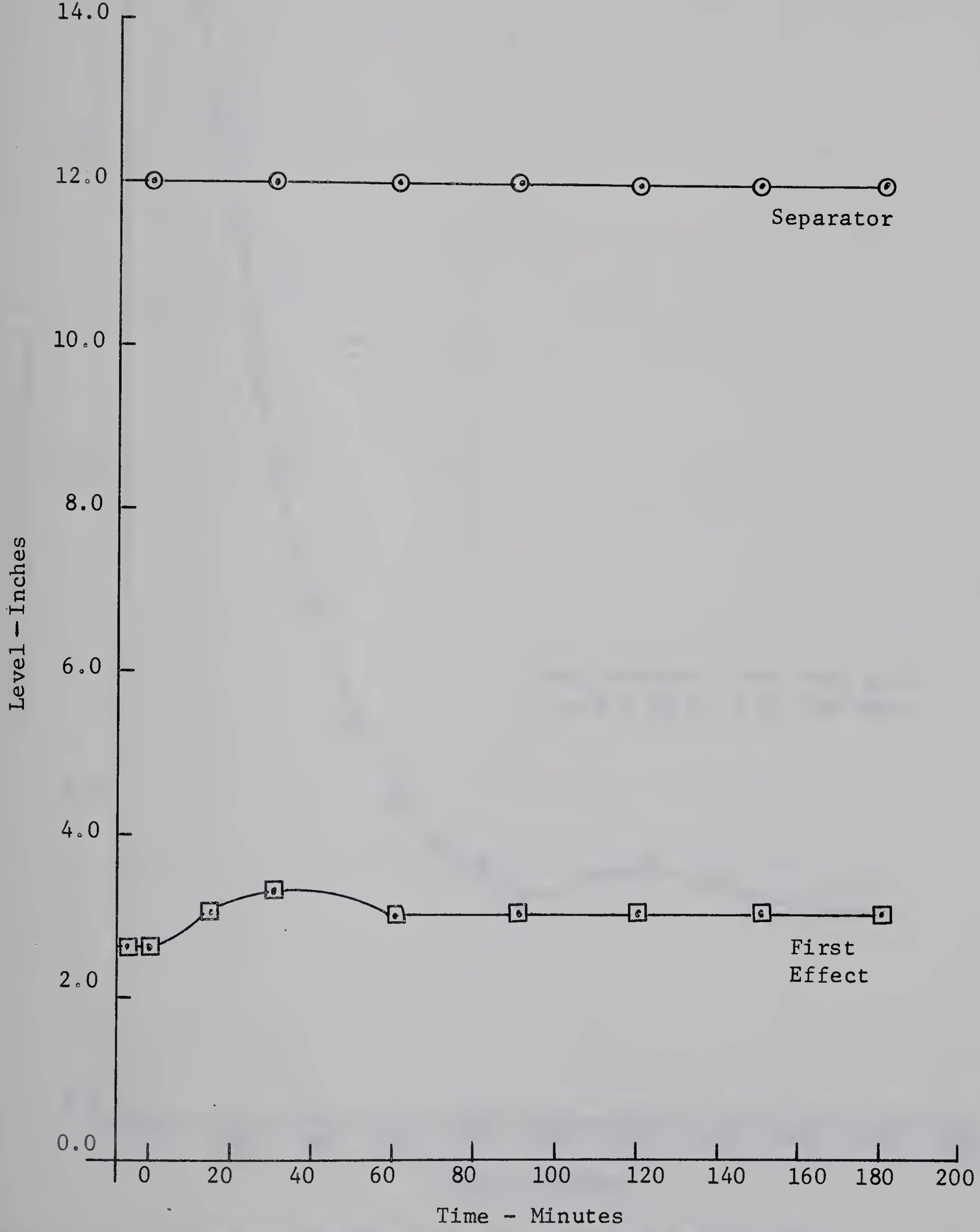


Figure 24 TRANSIENT RESPONSE OF SYSTEM LEVELS FOR RUN NO. 30L

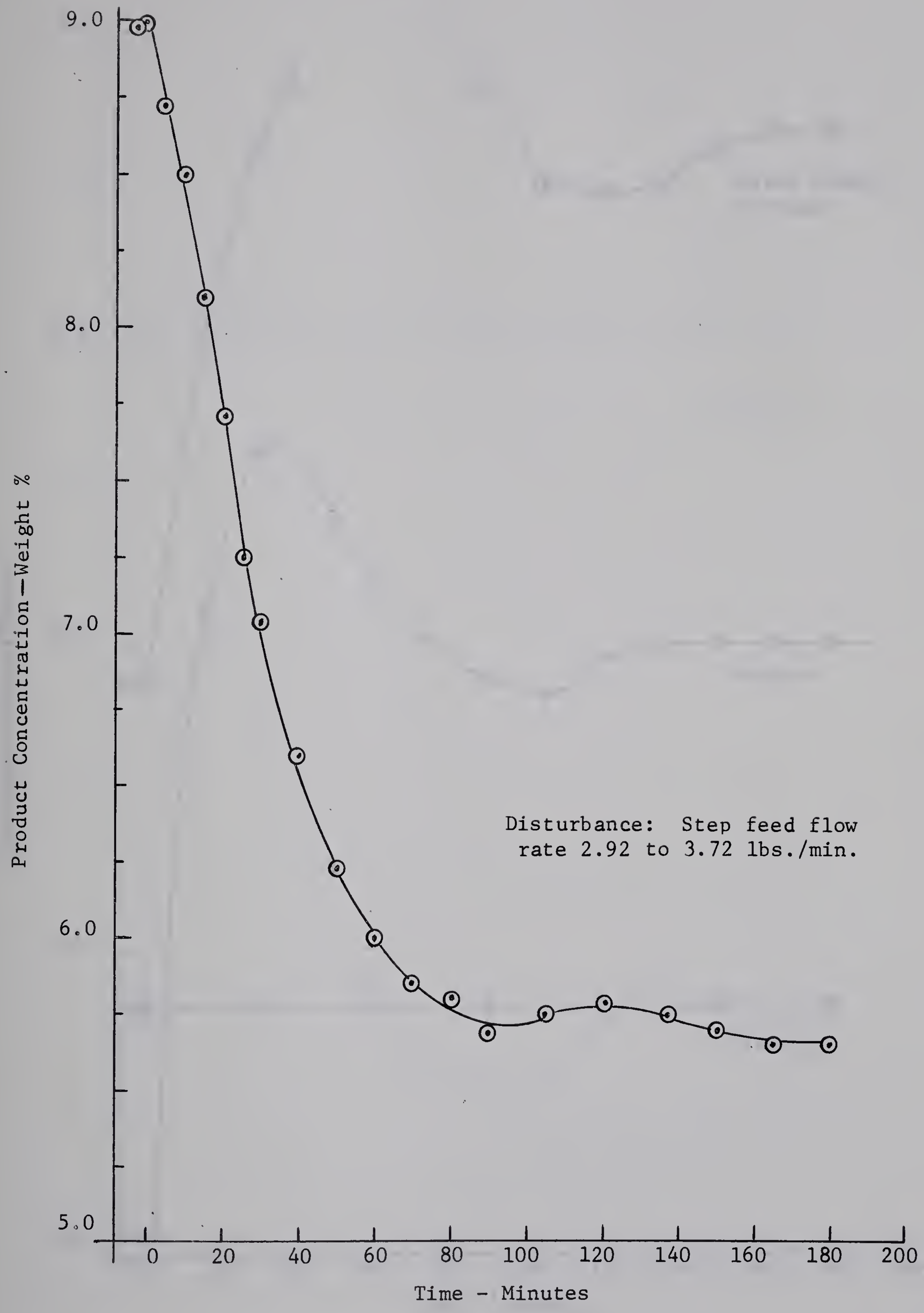
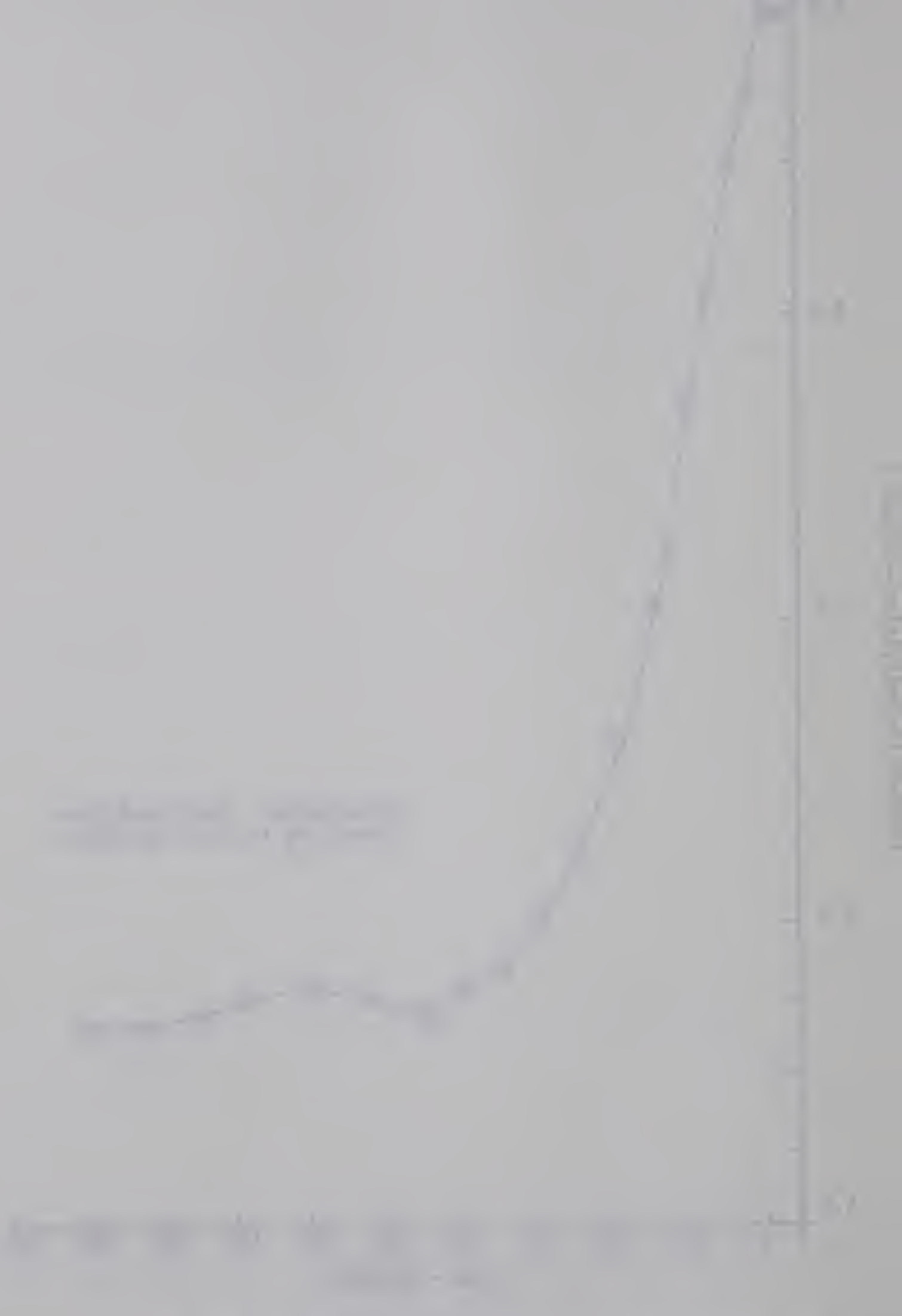


Figure 25 TRANSIENT CONCENTRATION RESPONSE FOR RUN NO. 40L



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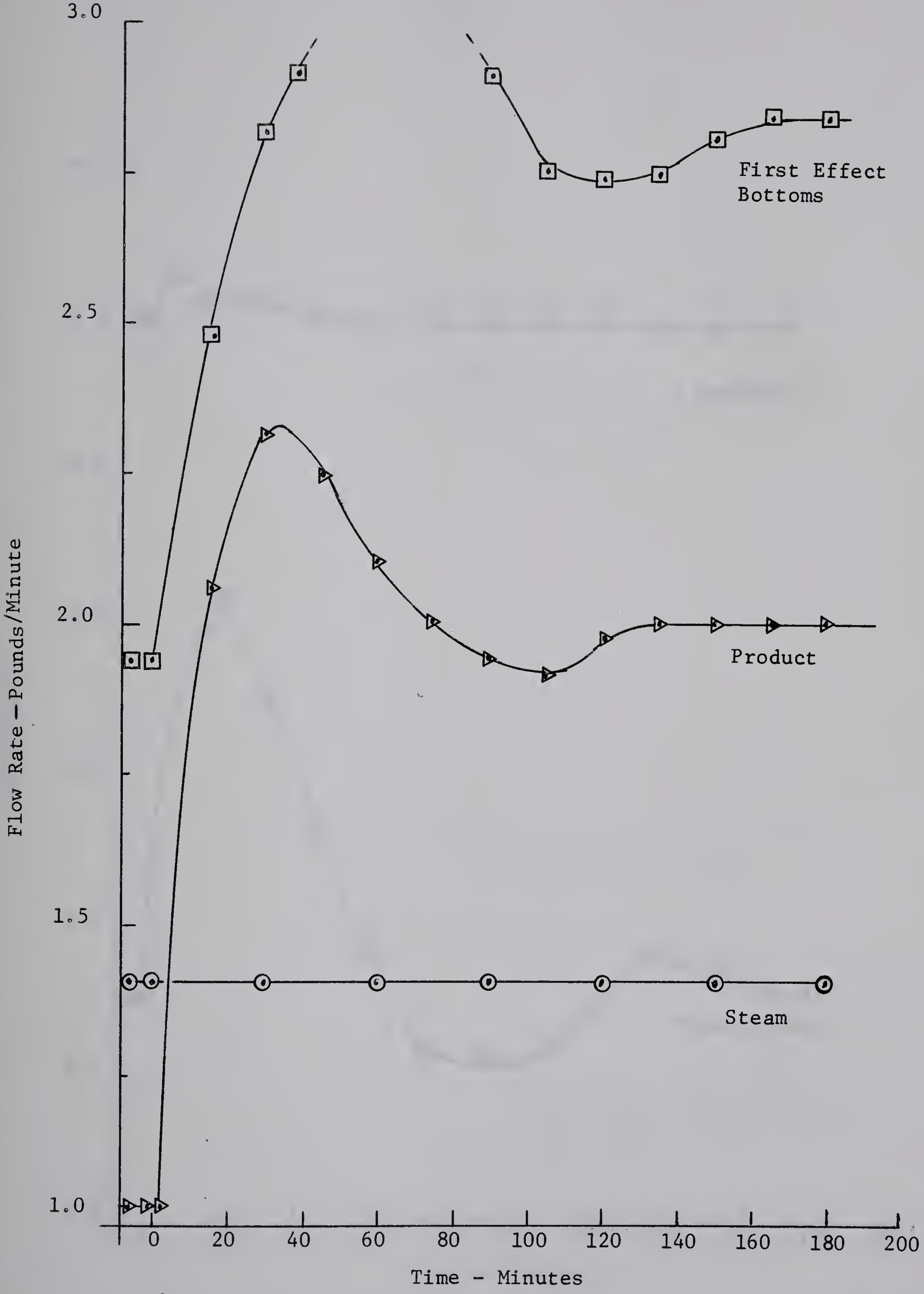
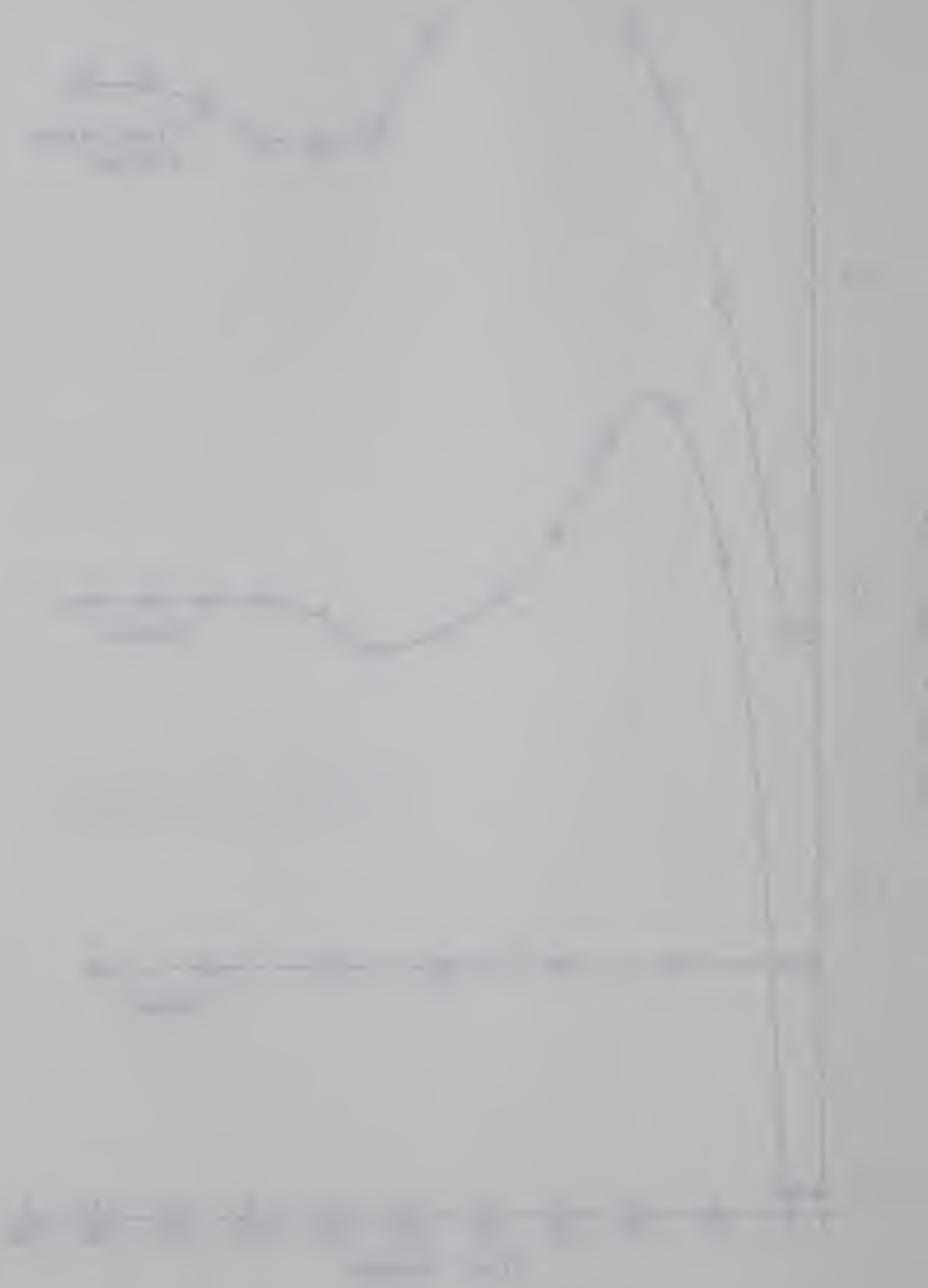


Figure 26 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN NO. 40L



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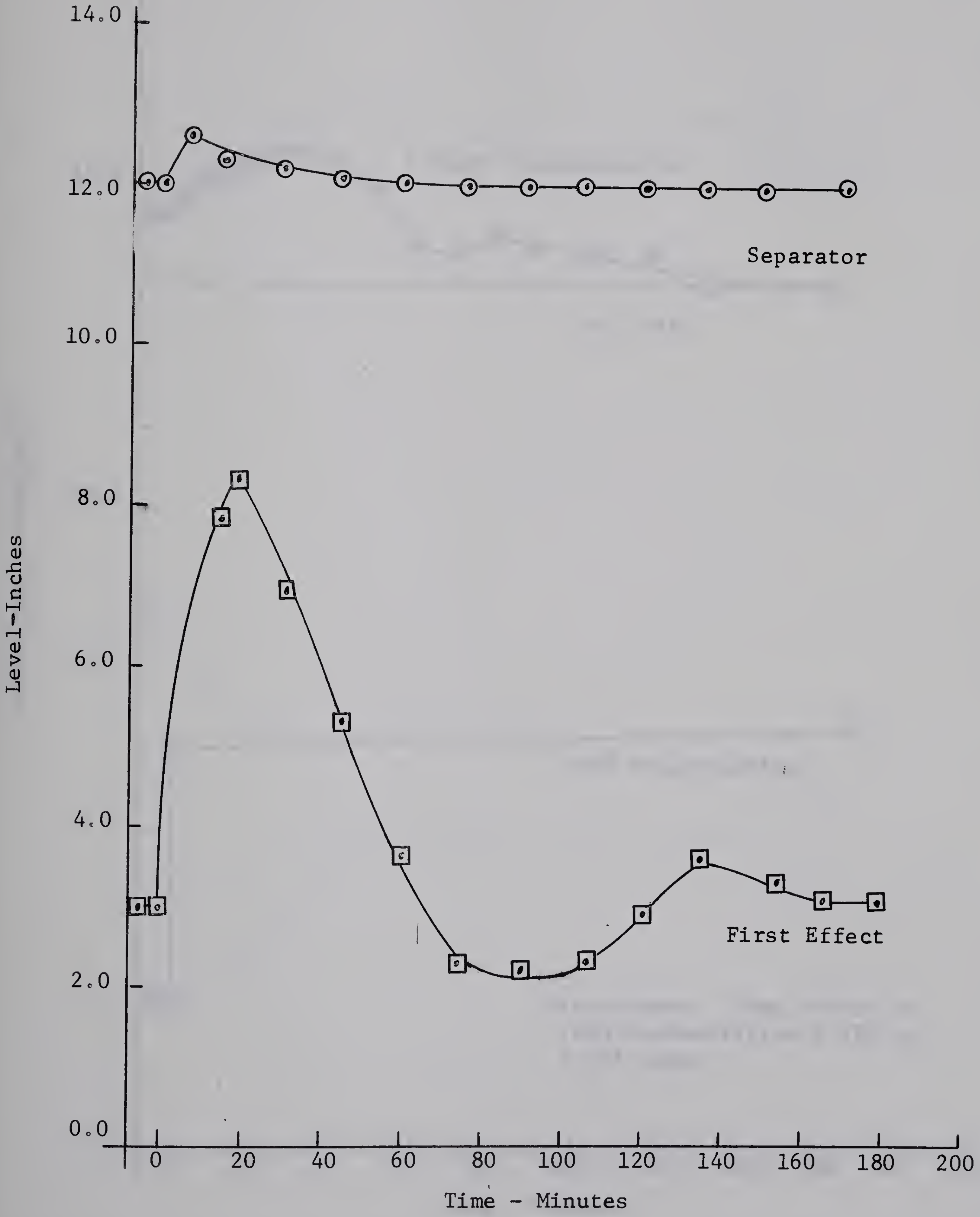


Figure 27 TRANSIENT RESPONSE OF SYSTEM LEVELS FOR RUN NO. 40L

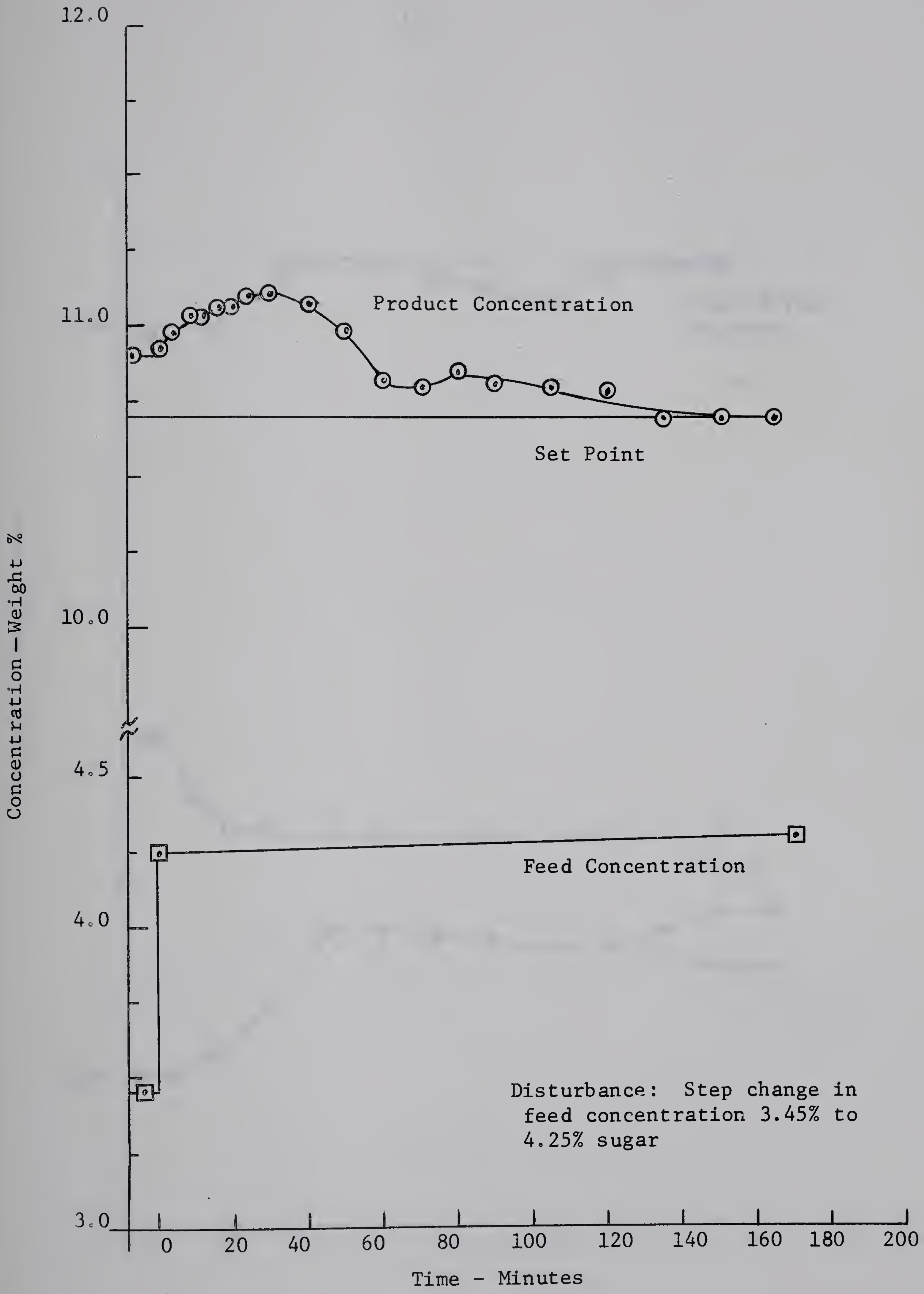


Figure 28 TRANSIENT CONCENTRATION RESPONSE FOR RUN NO. 8FF

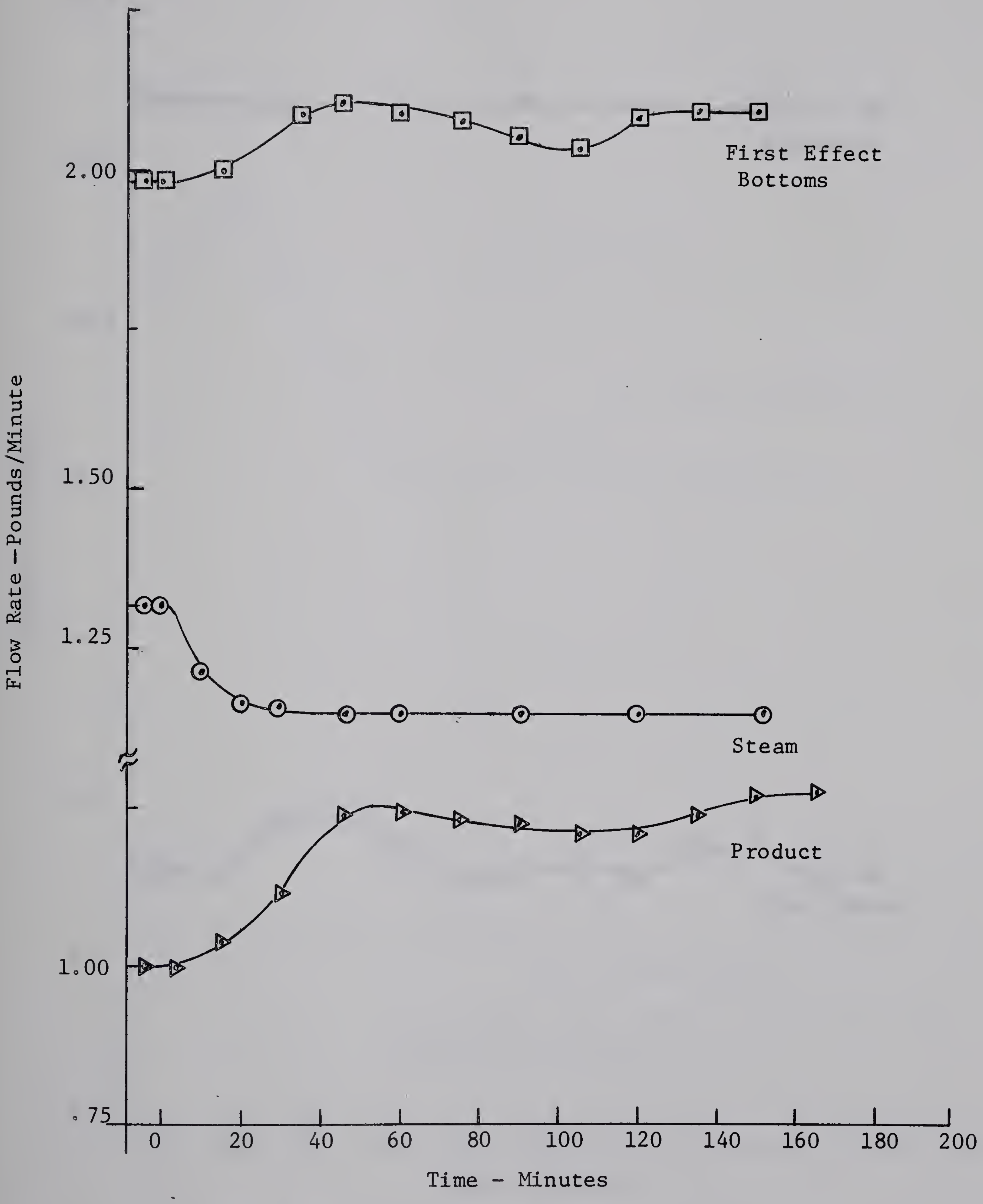


Figure 29 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN NO. 8FF

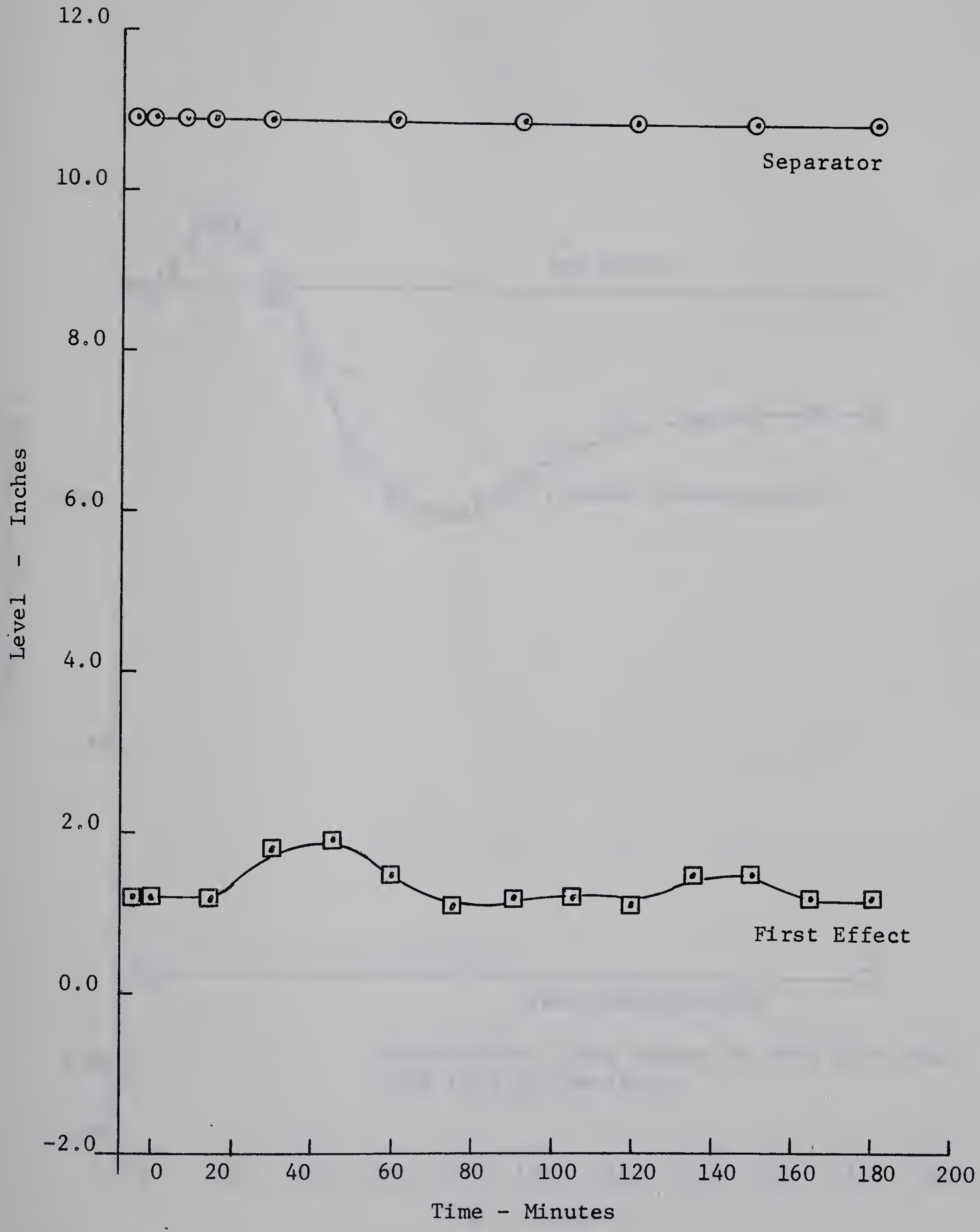
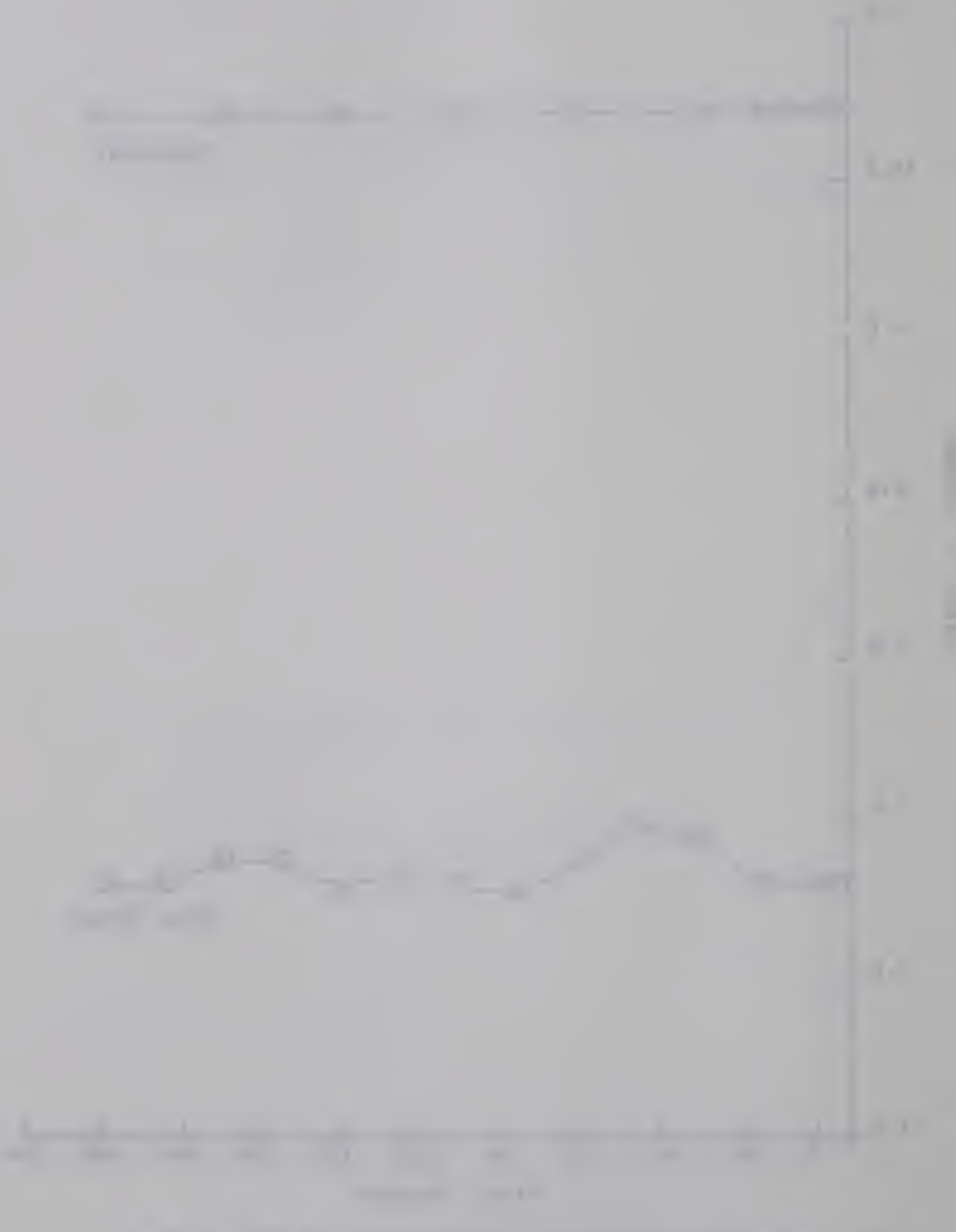


Figure 30 TRANSIENT RESPONSE OF SYSTEM LEVELS FOR RUN NO. 8FF



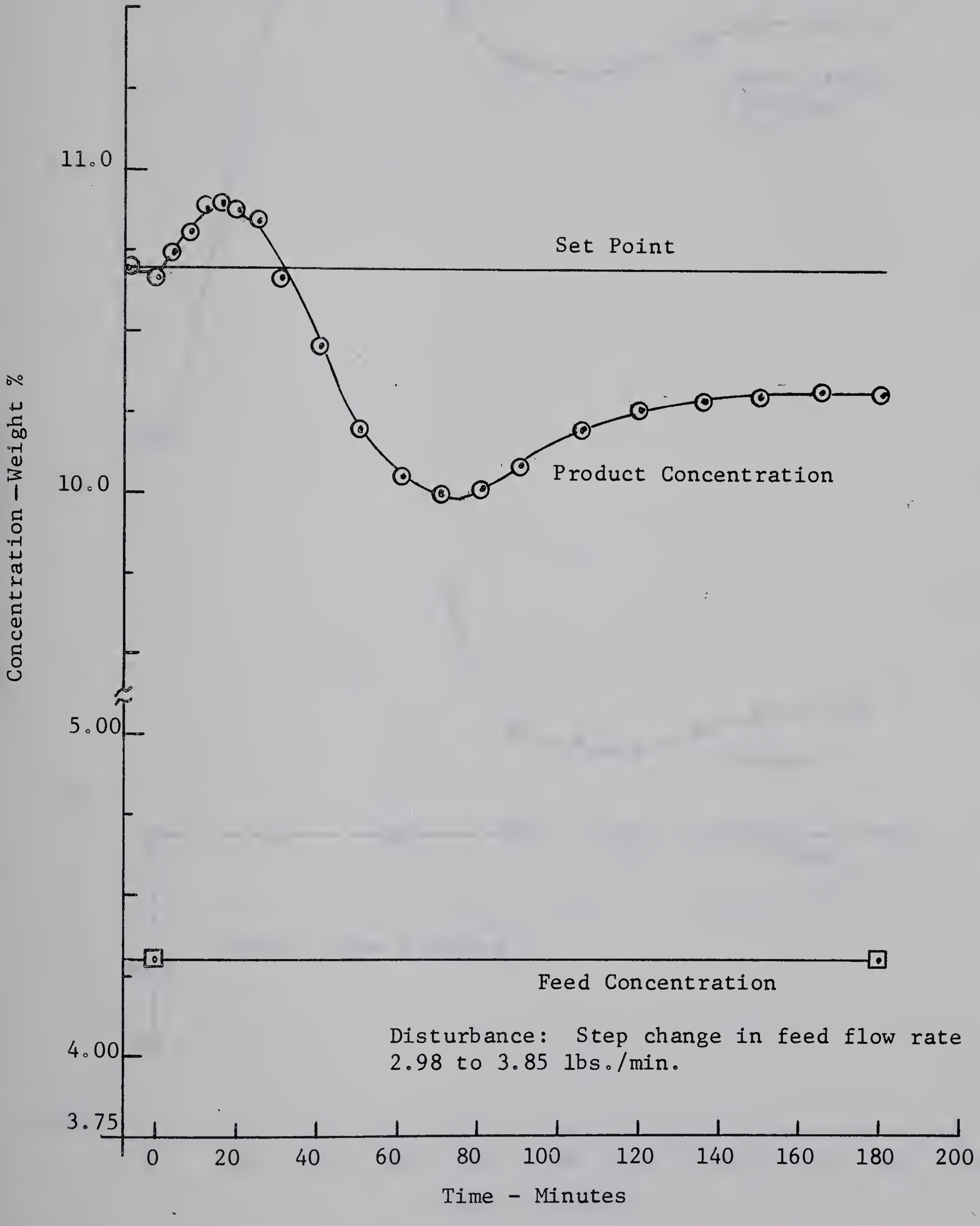


Figure 31 TRANSIENT CONCENTRATION RESPONSE FOR RUN NO. 9FF

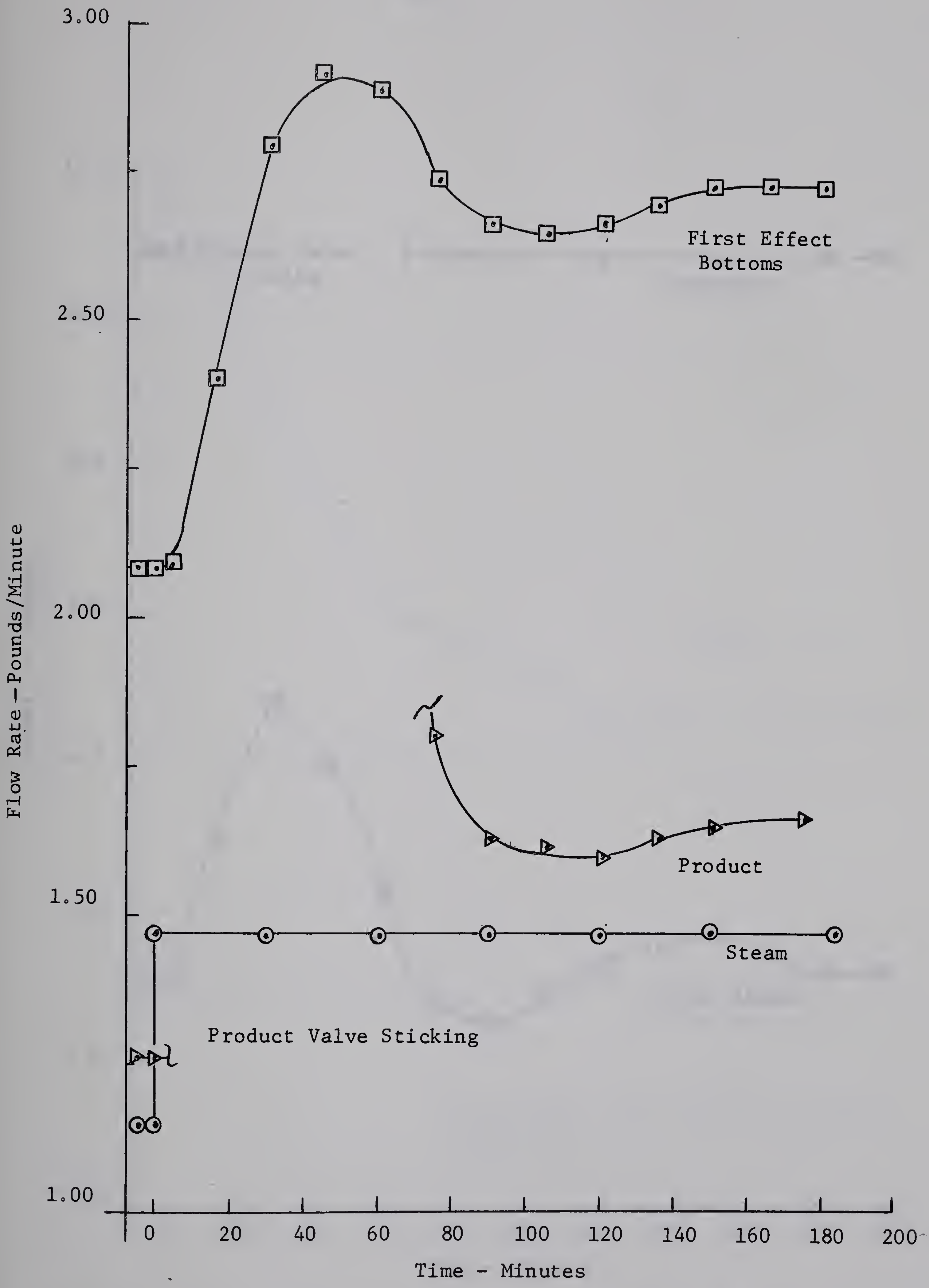
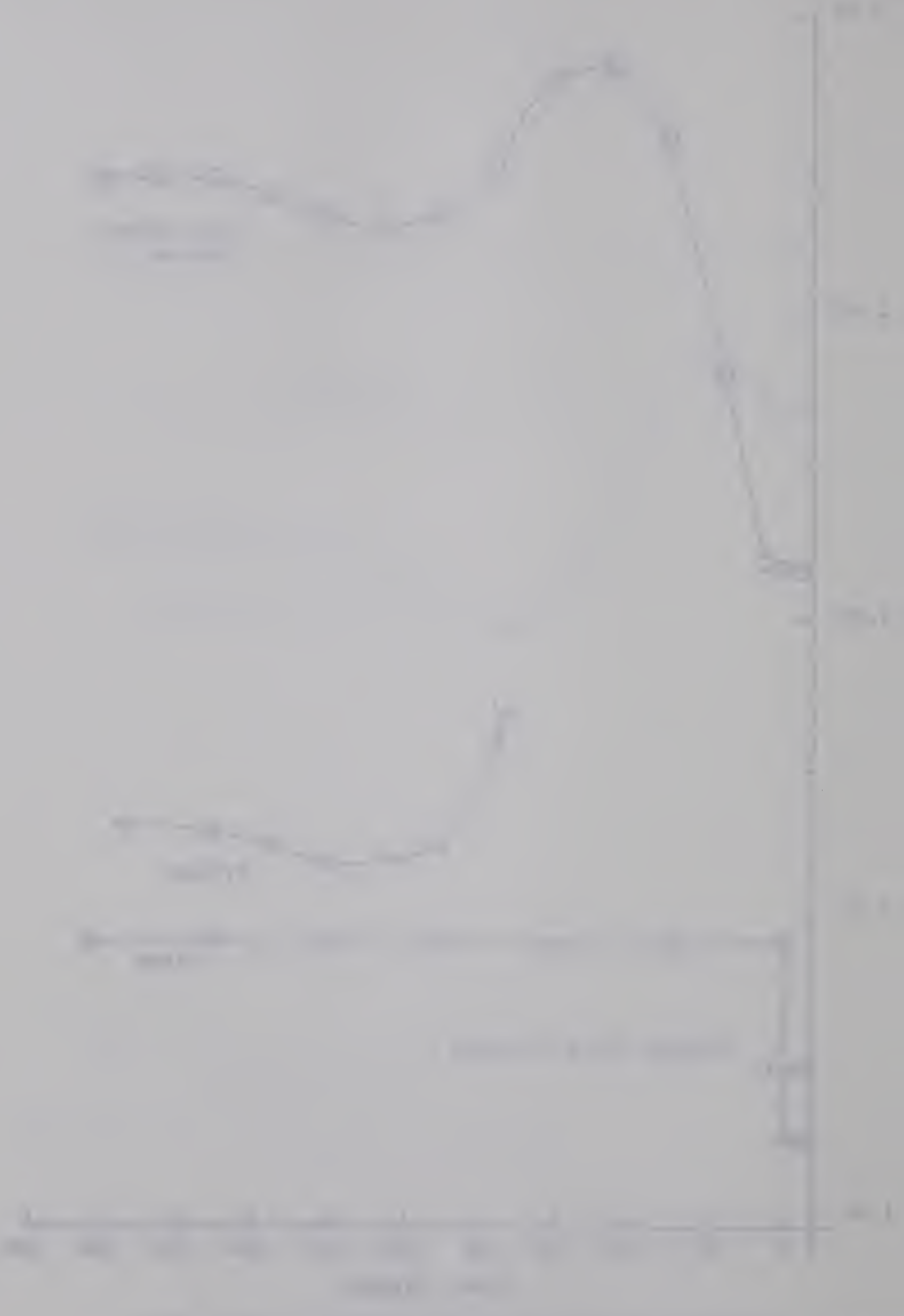


Figure 32 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN NO. 9FF



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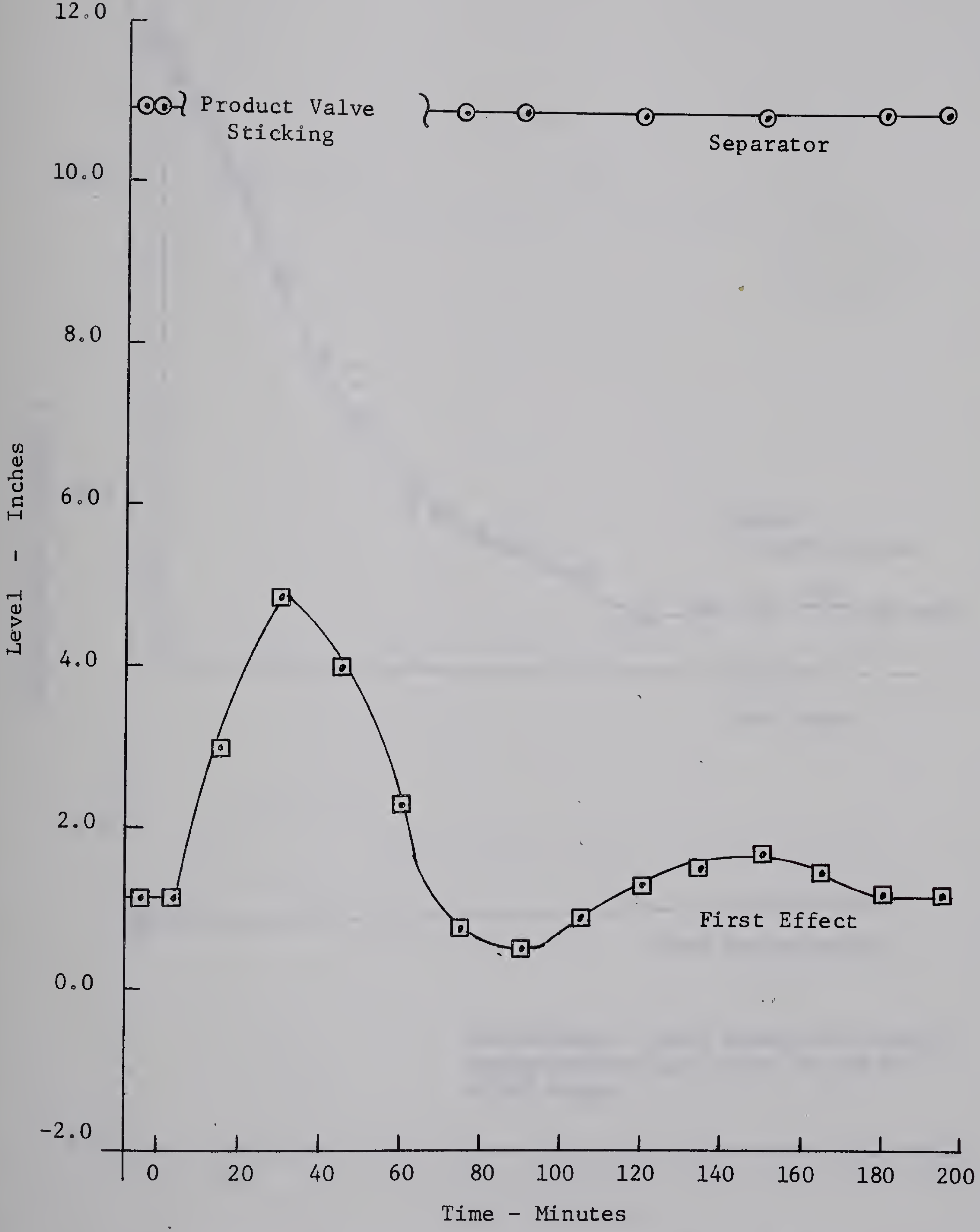


Figure 33 TRANSIENT RESPONSE OF SYSTEM LEVELS FOR RUN NO. 9FF

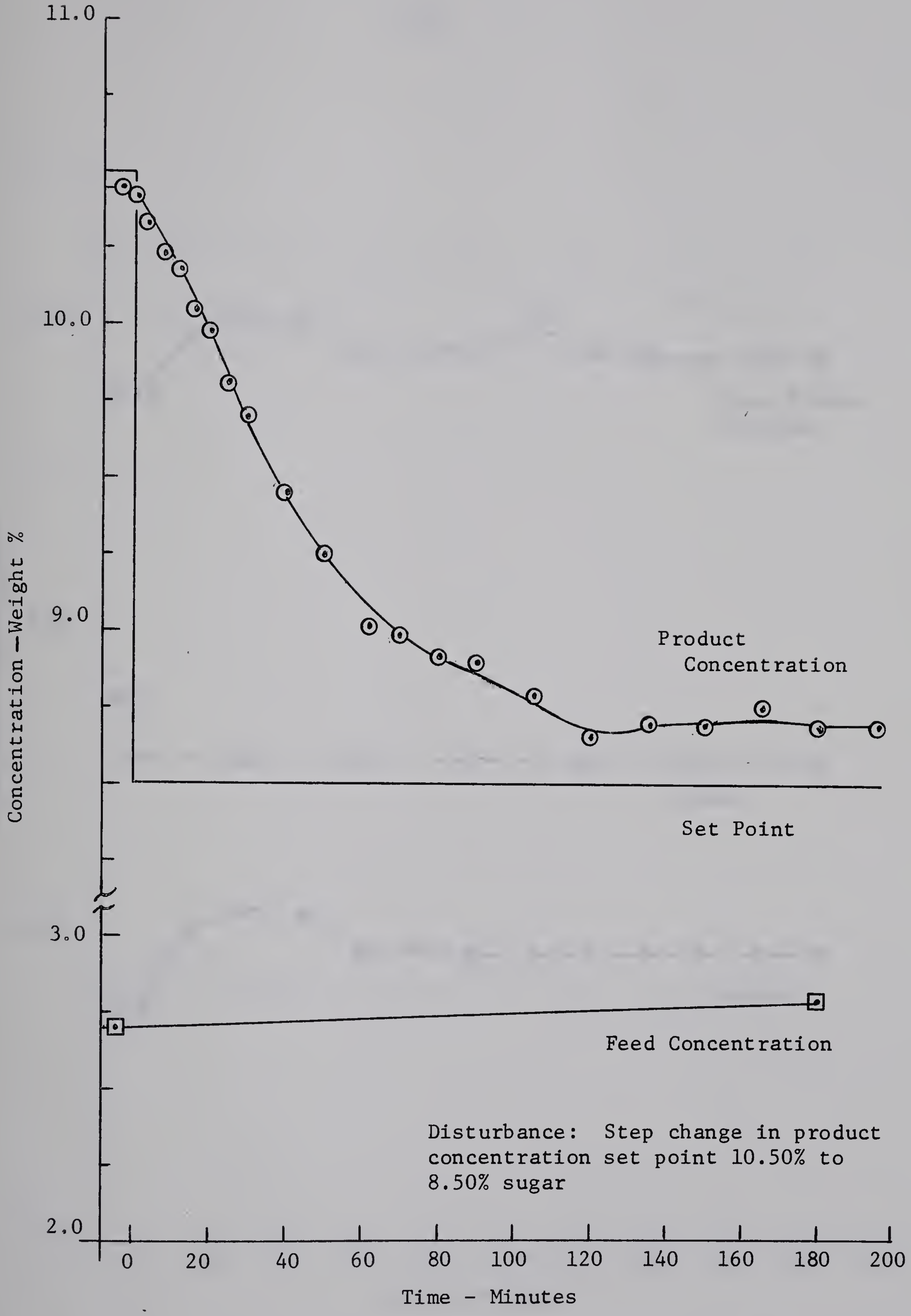


Figure 34 TRANSIENT CONCENTRATION RESPONSE FOR RUN NO. 10FF

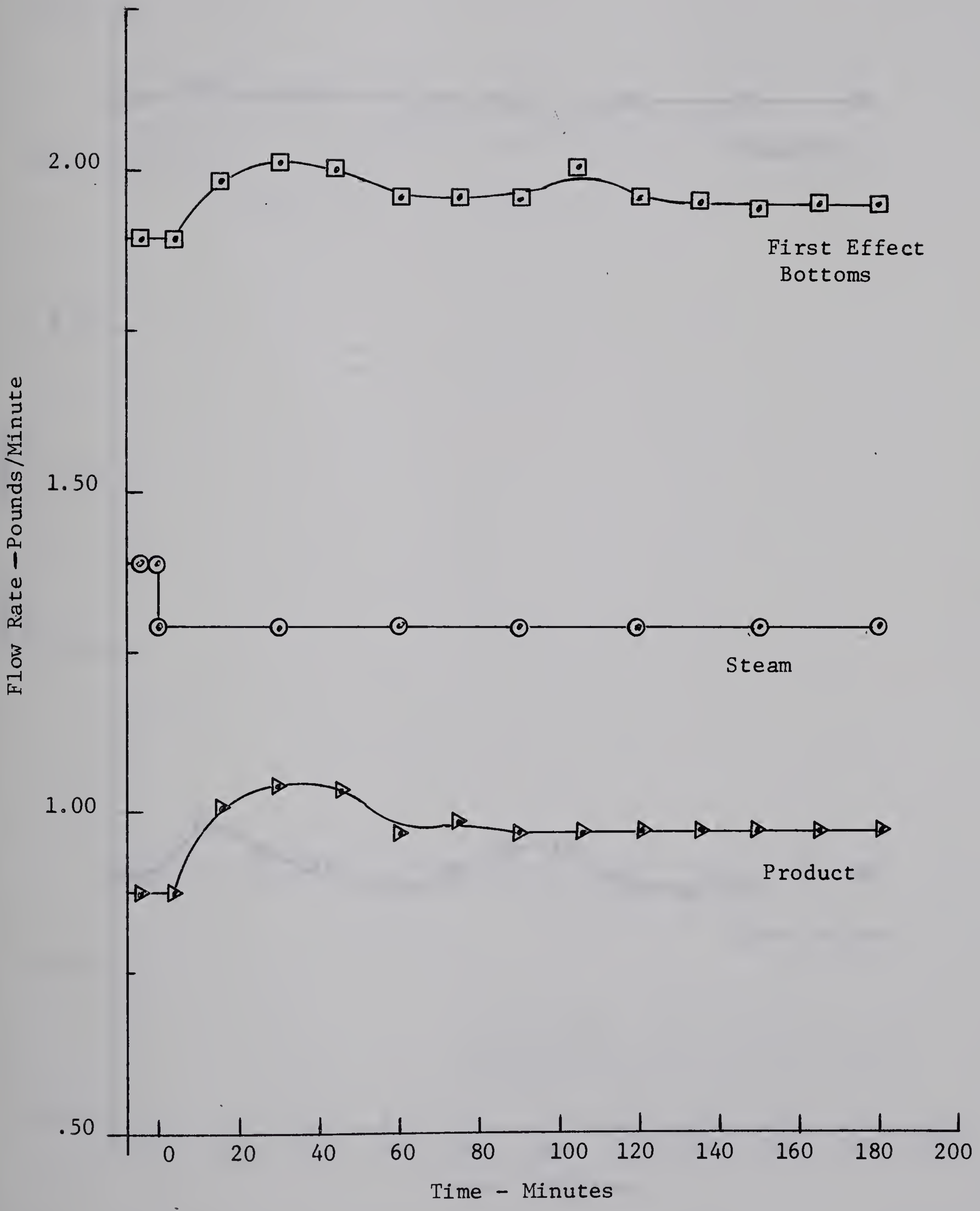
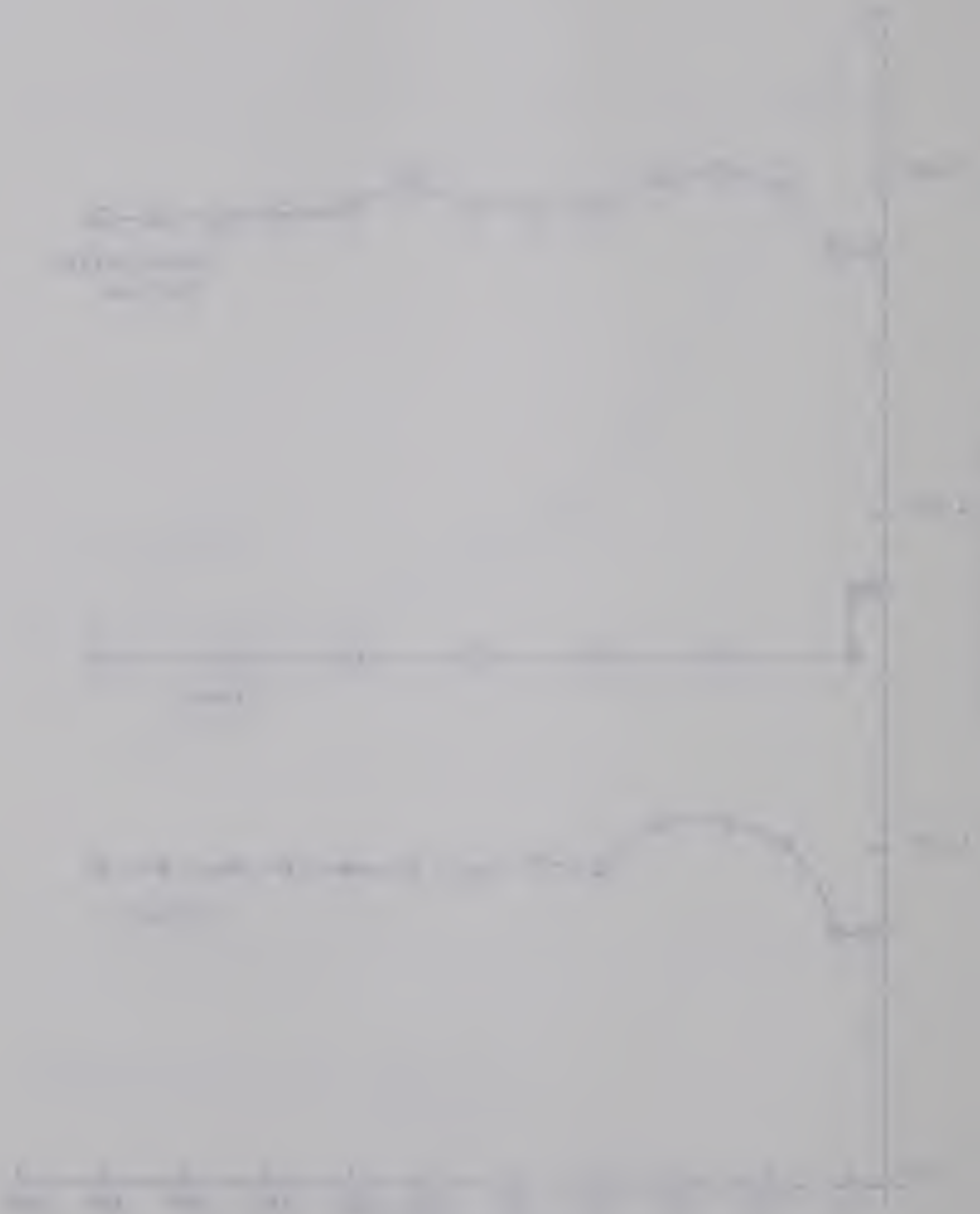


Figure 35 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN NO. 10FF



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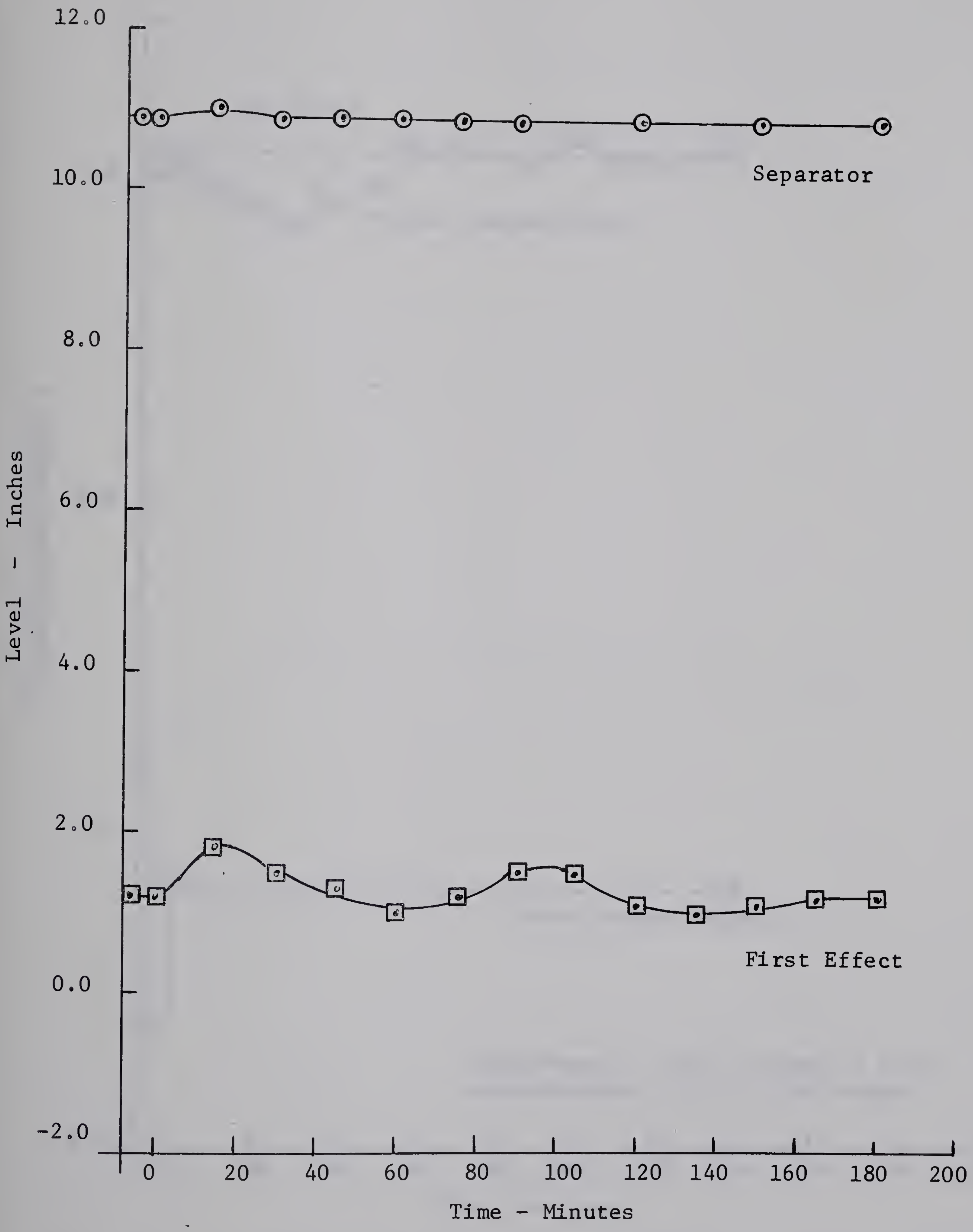


Figure 36 TRANSIENT RESPONSE OF SYSTEM LEVELS FOR RUN NO. 10FF

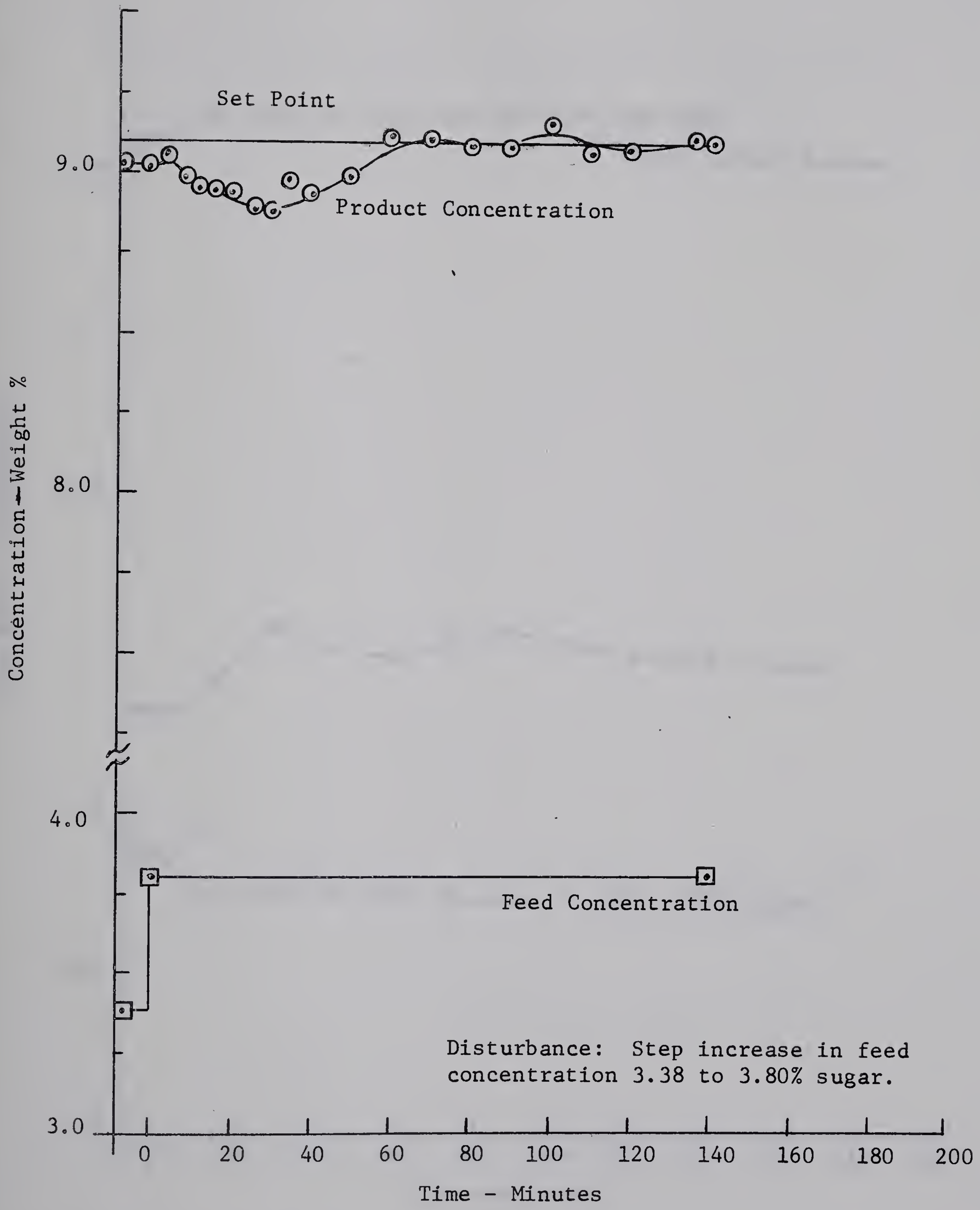


Figure 37 TRANSIENT CONCENTRATION RESPONSE FOR RUN NO. 12FFB

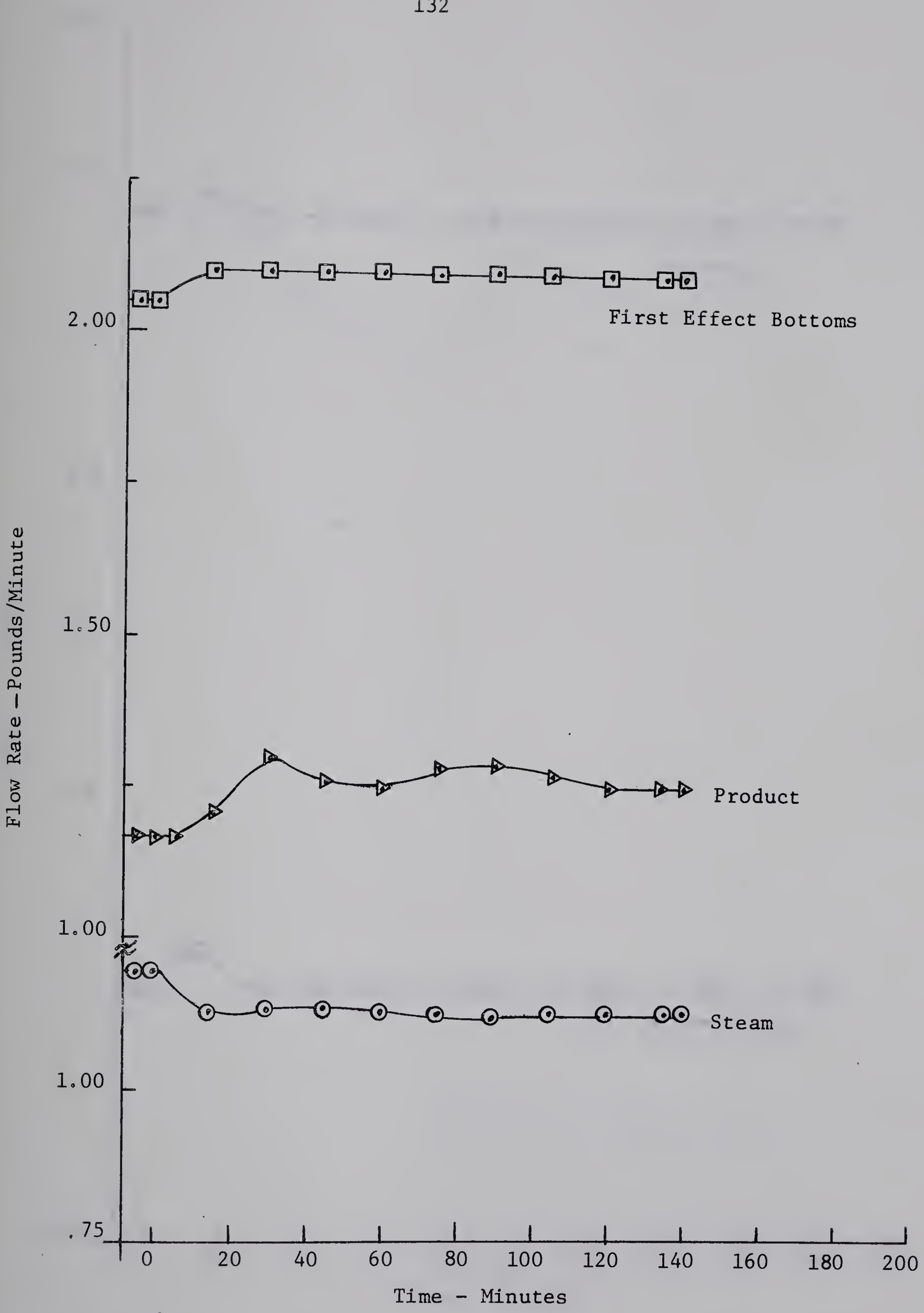
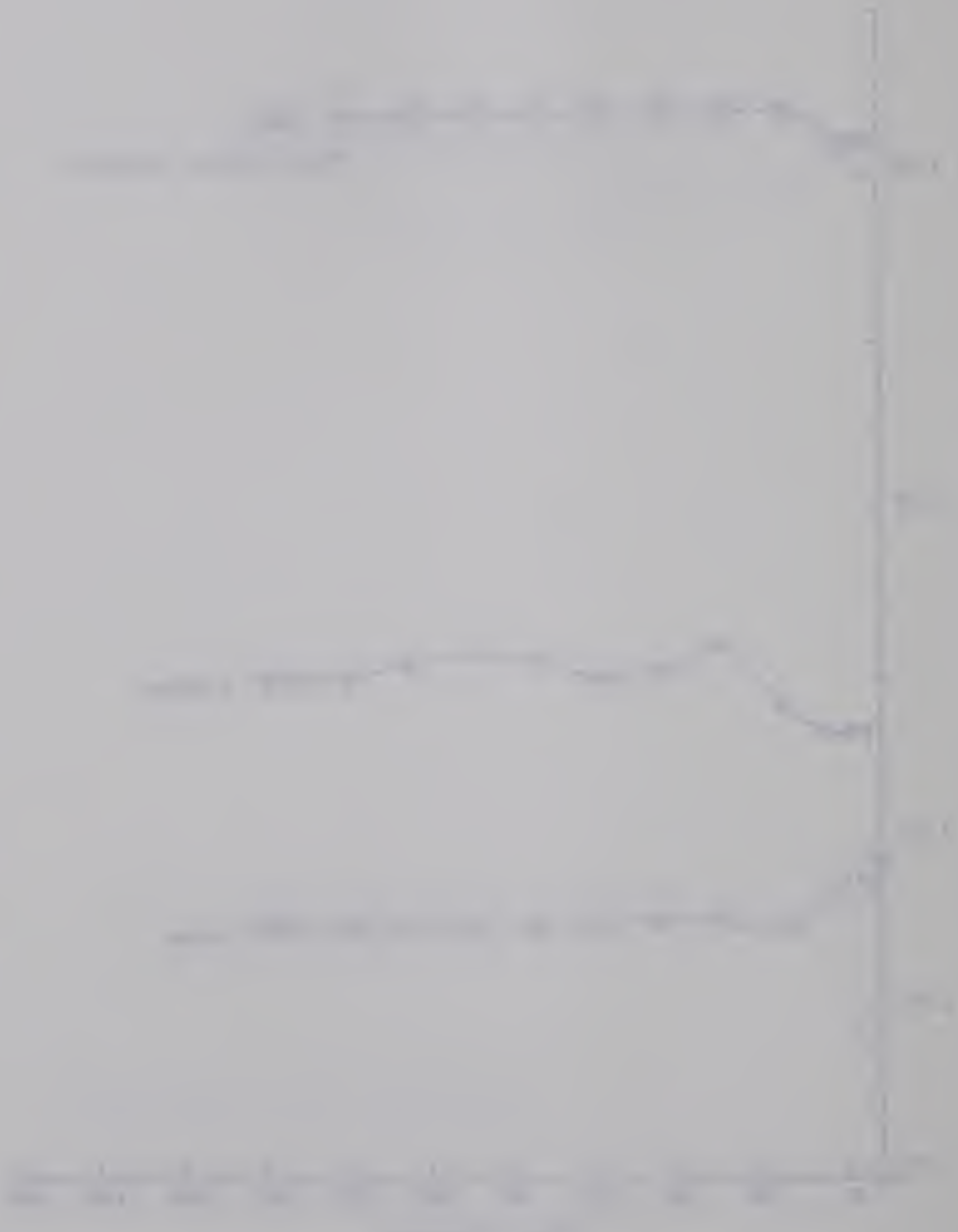


Figure 38 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN NO. 12FFB



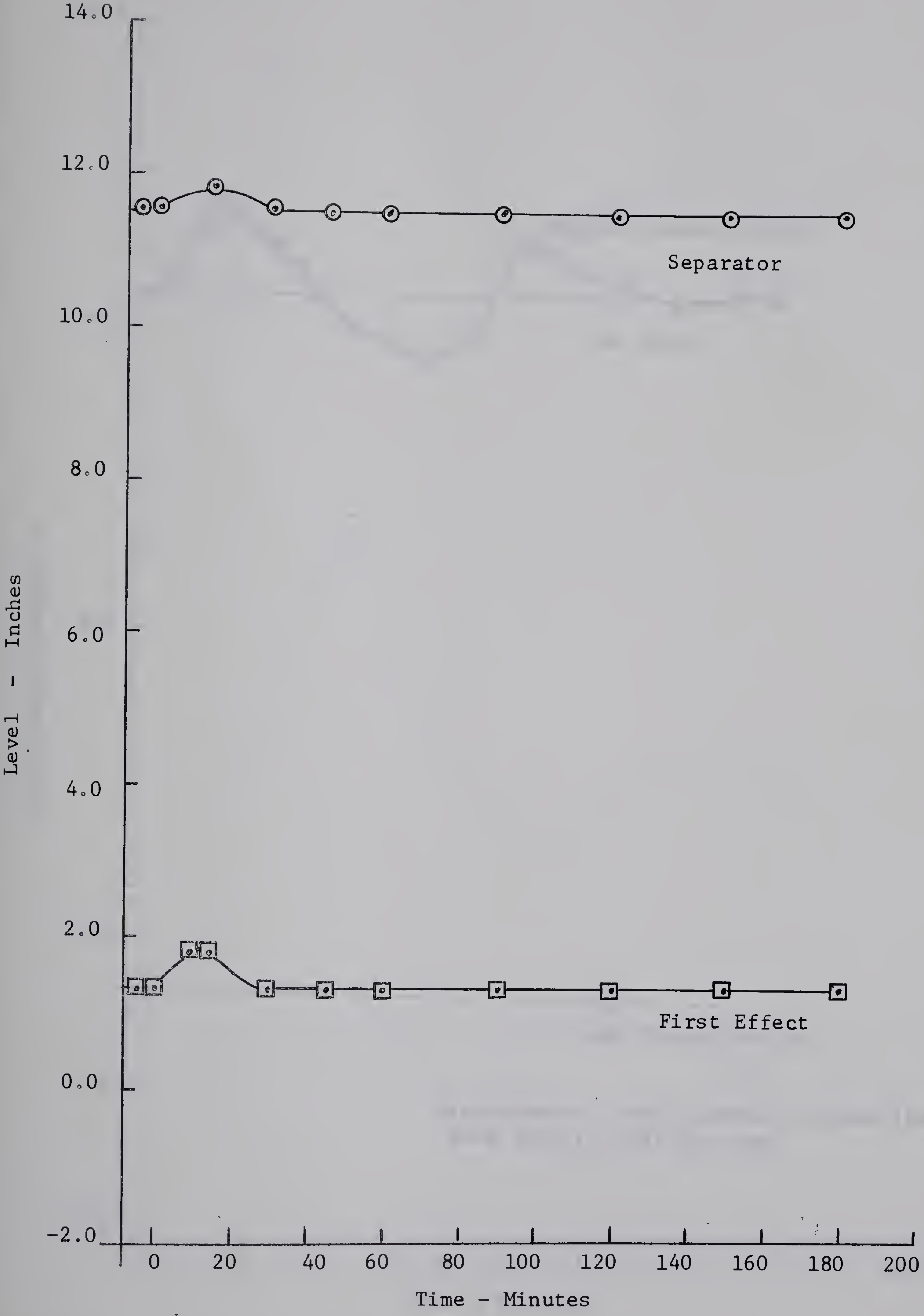


Figure 39 TRANSIENT RESPONSE OF SYSTEM LEVELS FOR RUN NO. 12FFB

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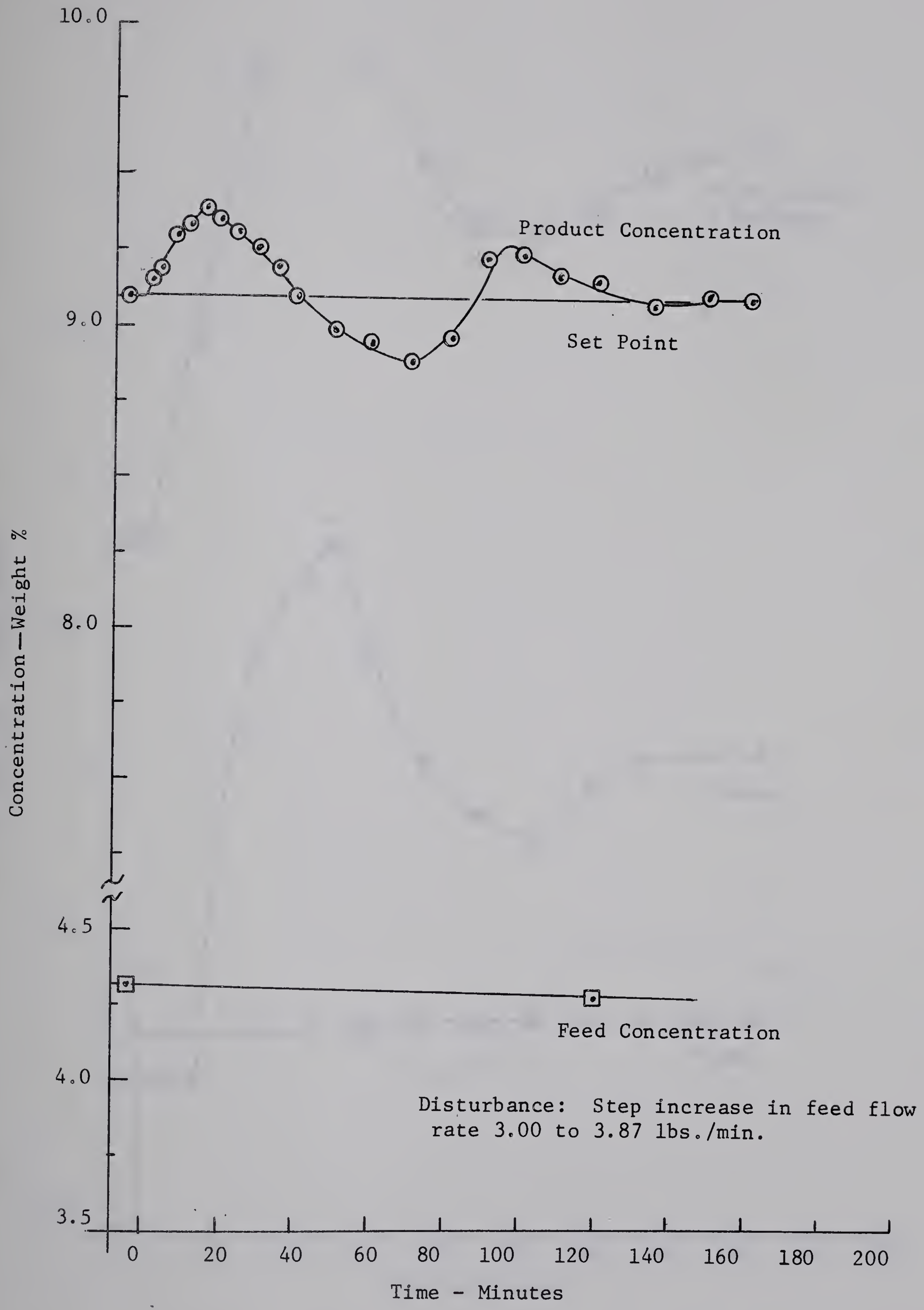


Figure 40 TRANSIENT CONCENTRATION RESPONSE FOR RUN NO. 15FFB

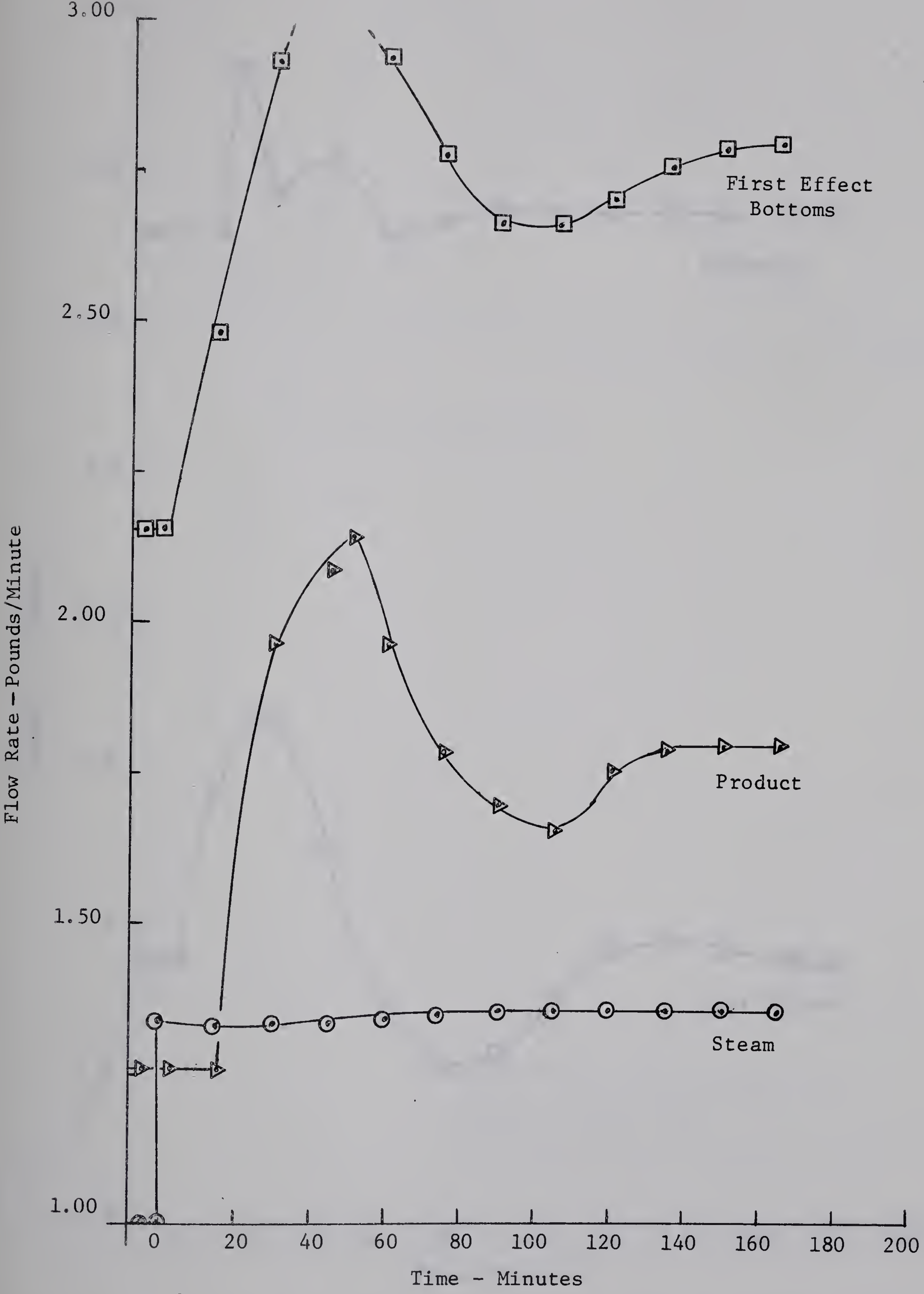


Figure 41 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN NO. 15FFB

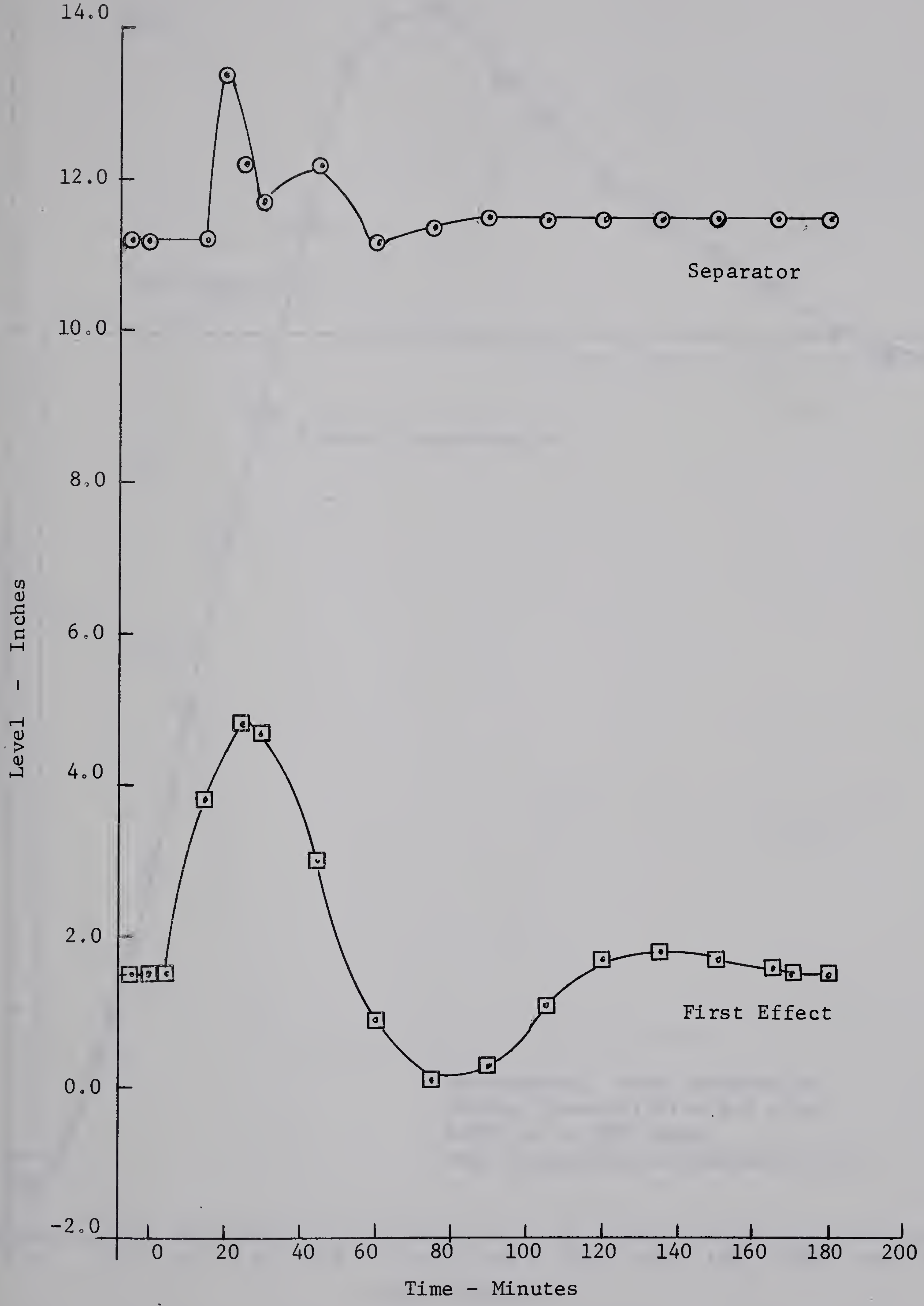
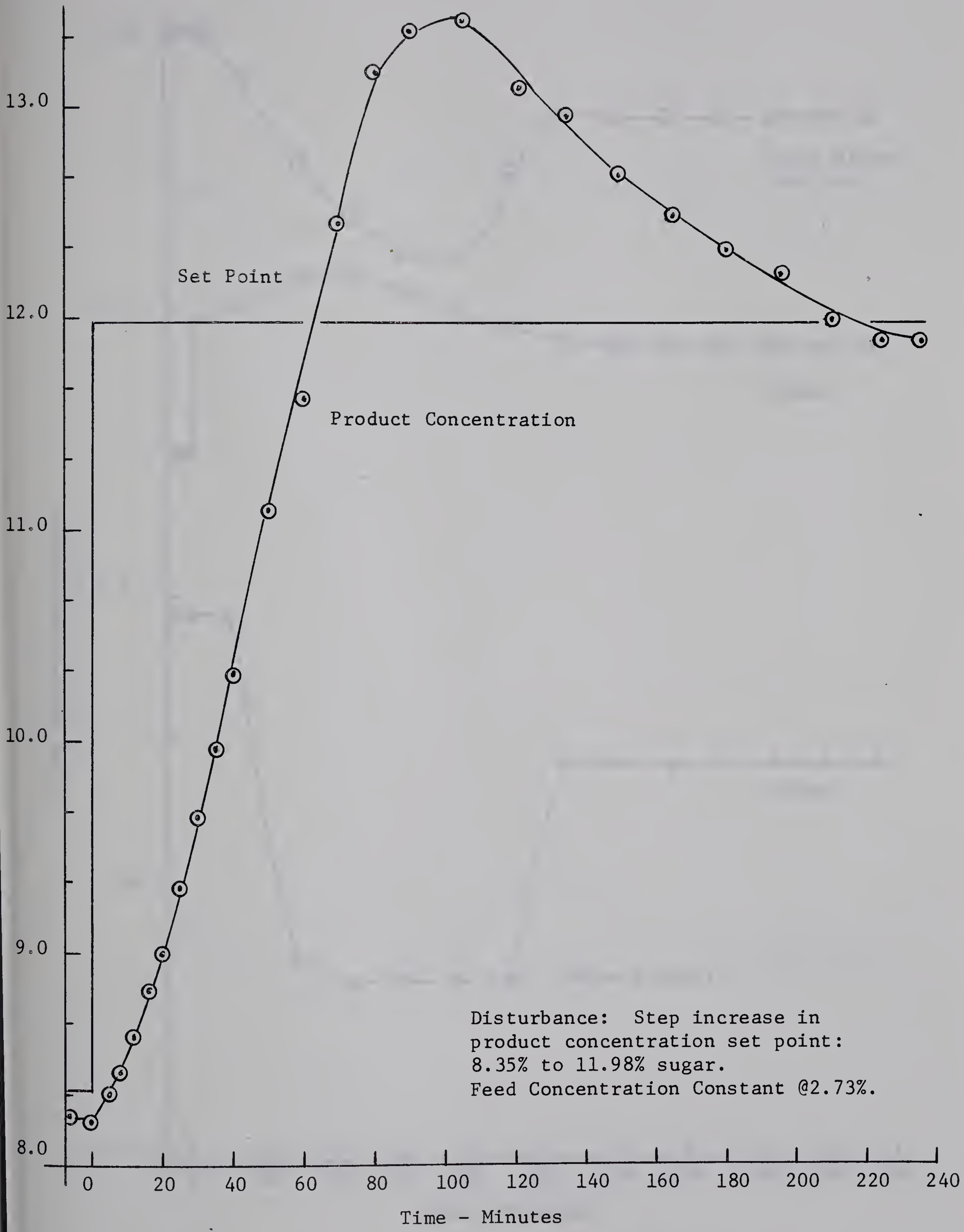


Figure 42 TRANSIENT RESPONSE OF SYSTEM LEVELS FOR RUN NO. 15FFB



Disturbance: Step increase in product concentration set point: 8.35% to 11.98% sugar.
Feed Concentration Constant @2.73%.

Figure 43 TRANSIENT CONCENTRATION RESPONSE FOR RUN NO. 16FFB

[illegible]

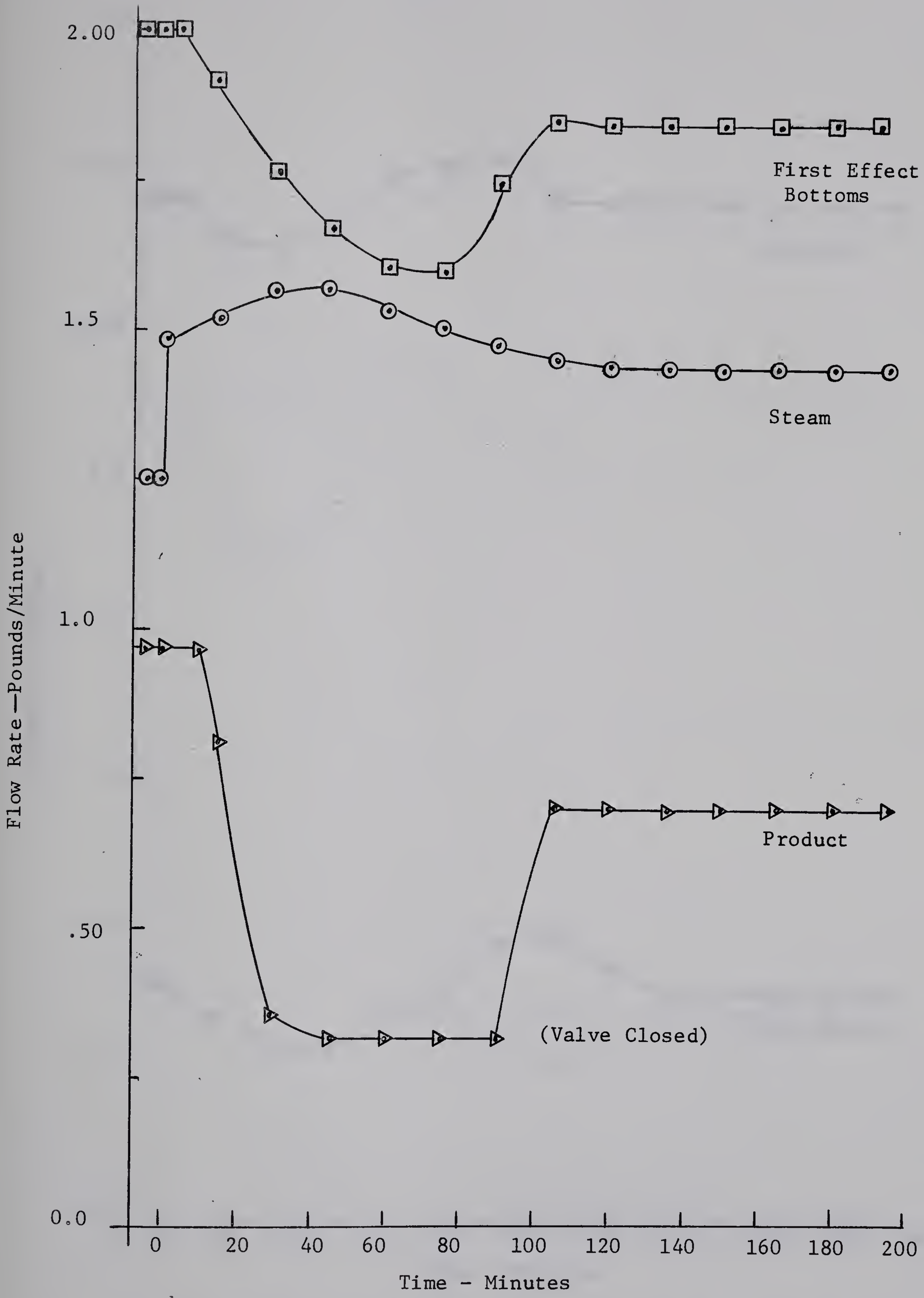


Figure 44 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN NO. 16FFB

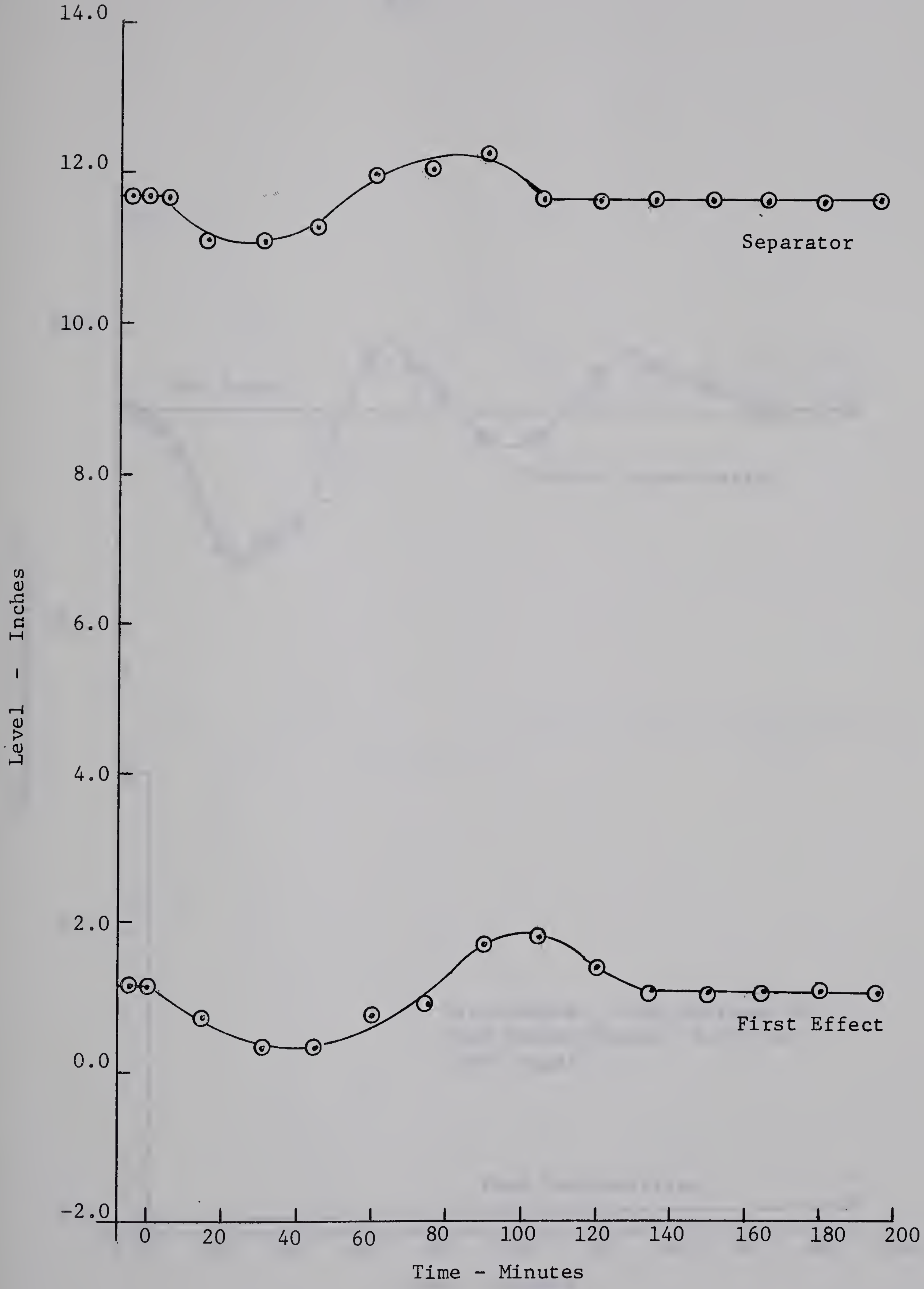


Figure 45 TRANSIENT RESPONSE OF SYSTEM LEVELS FOR RUN NO. 16FFB

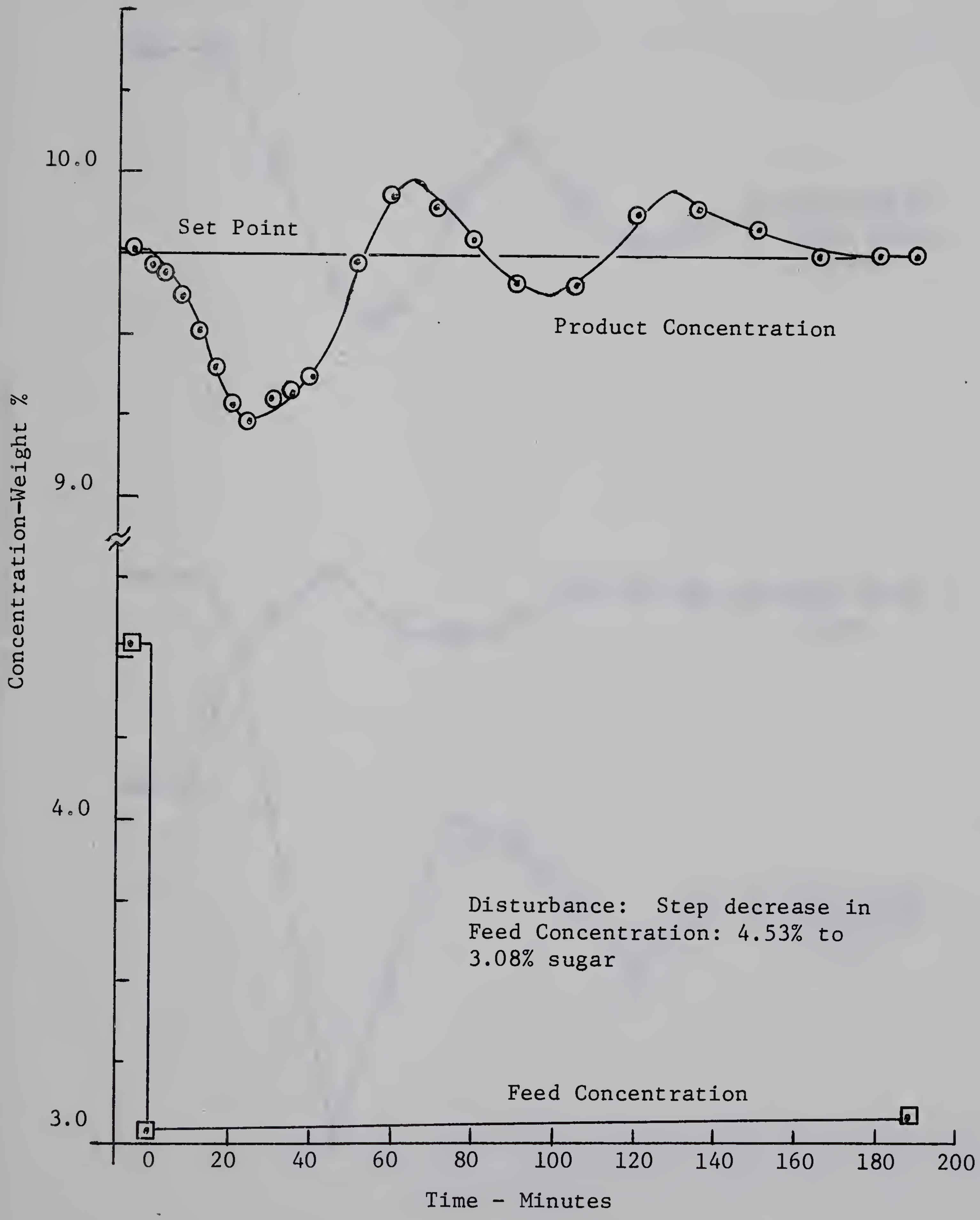
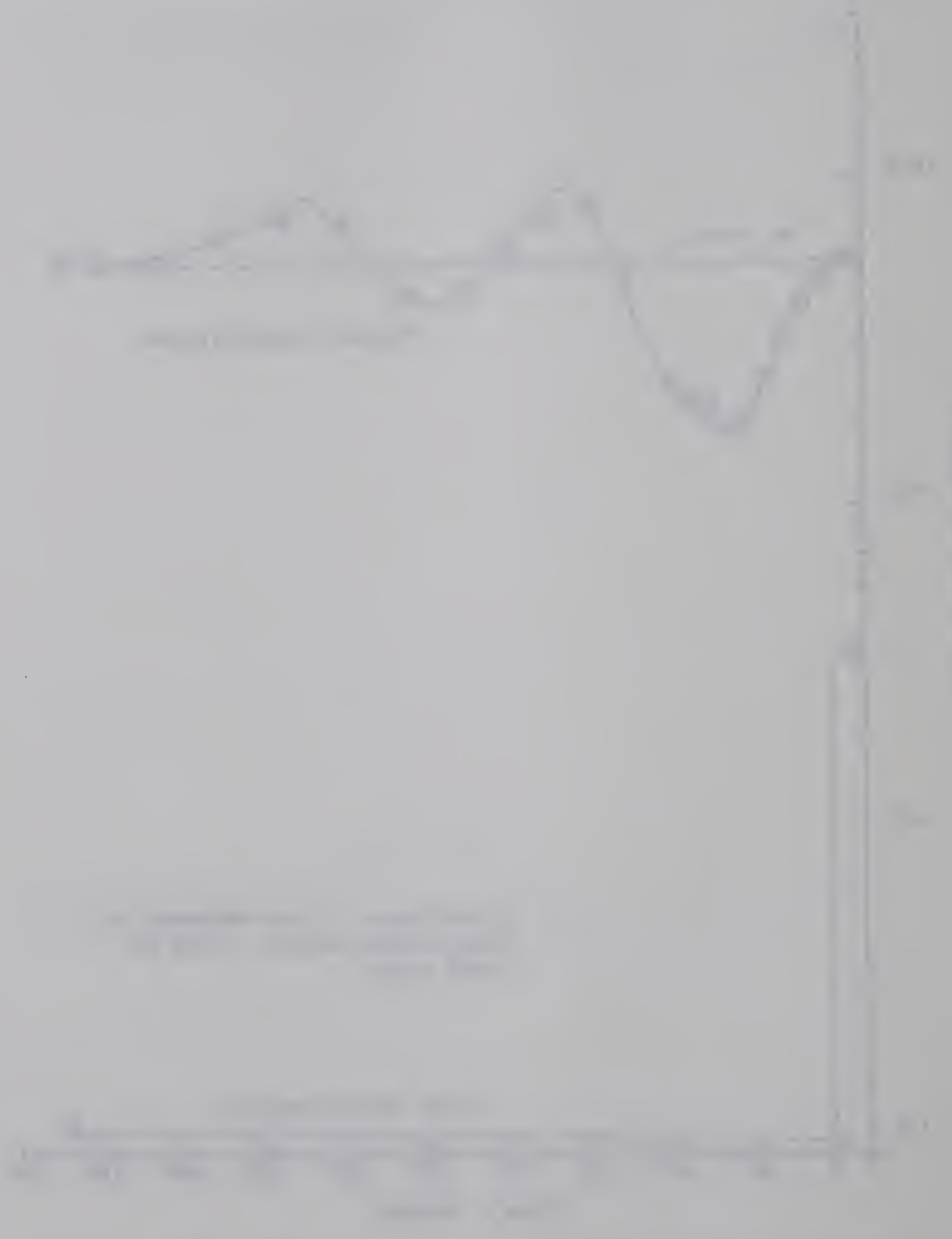


Figure 46 TRANSIENT CONCENTRATION RESPONSE FOR RUN NO. 17FB



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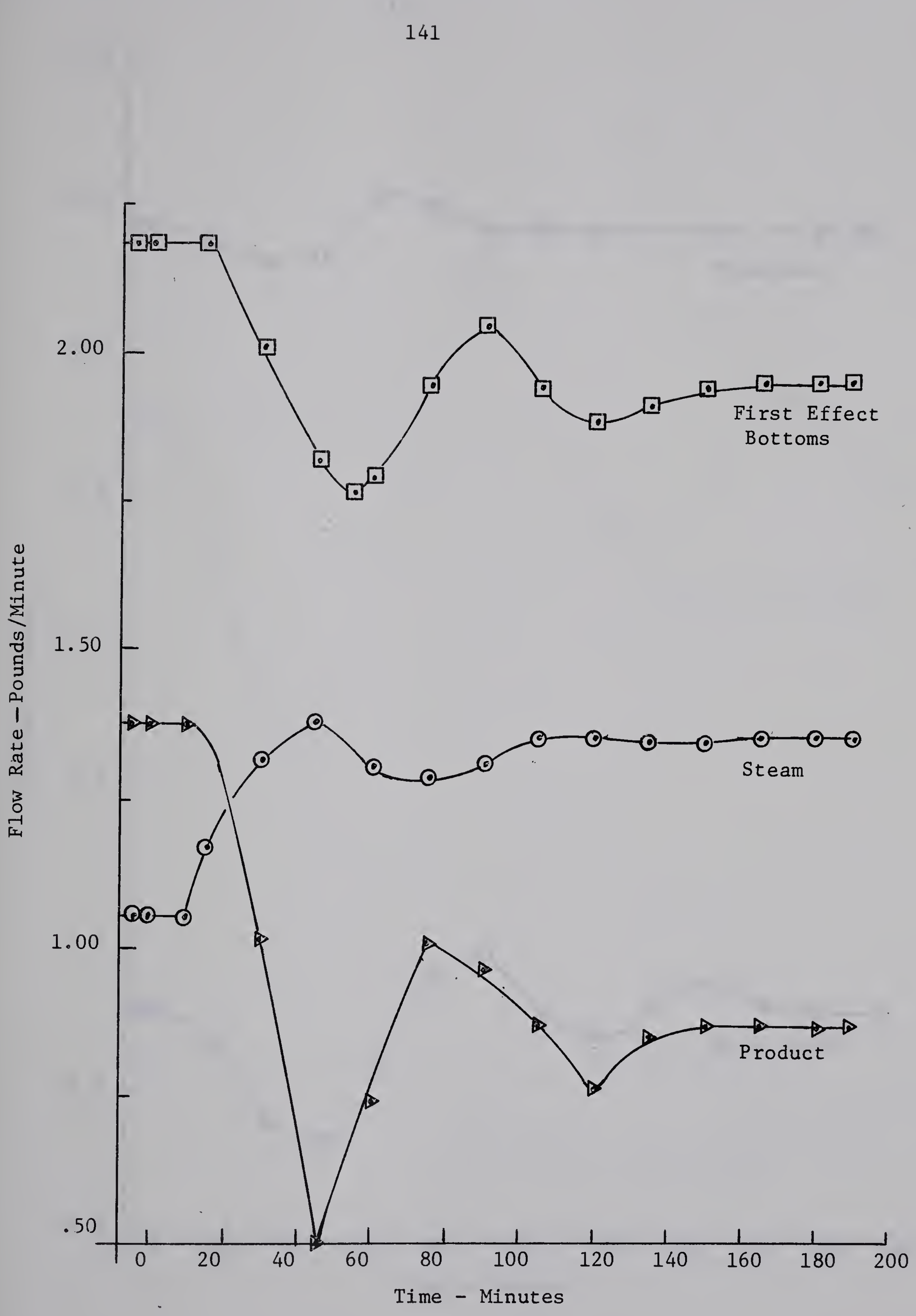
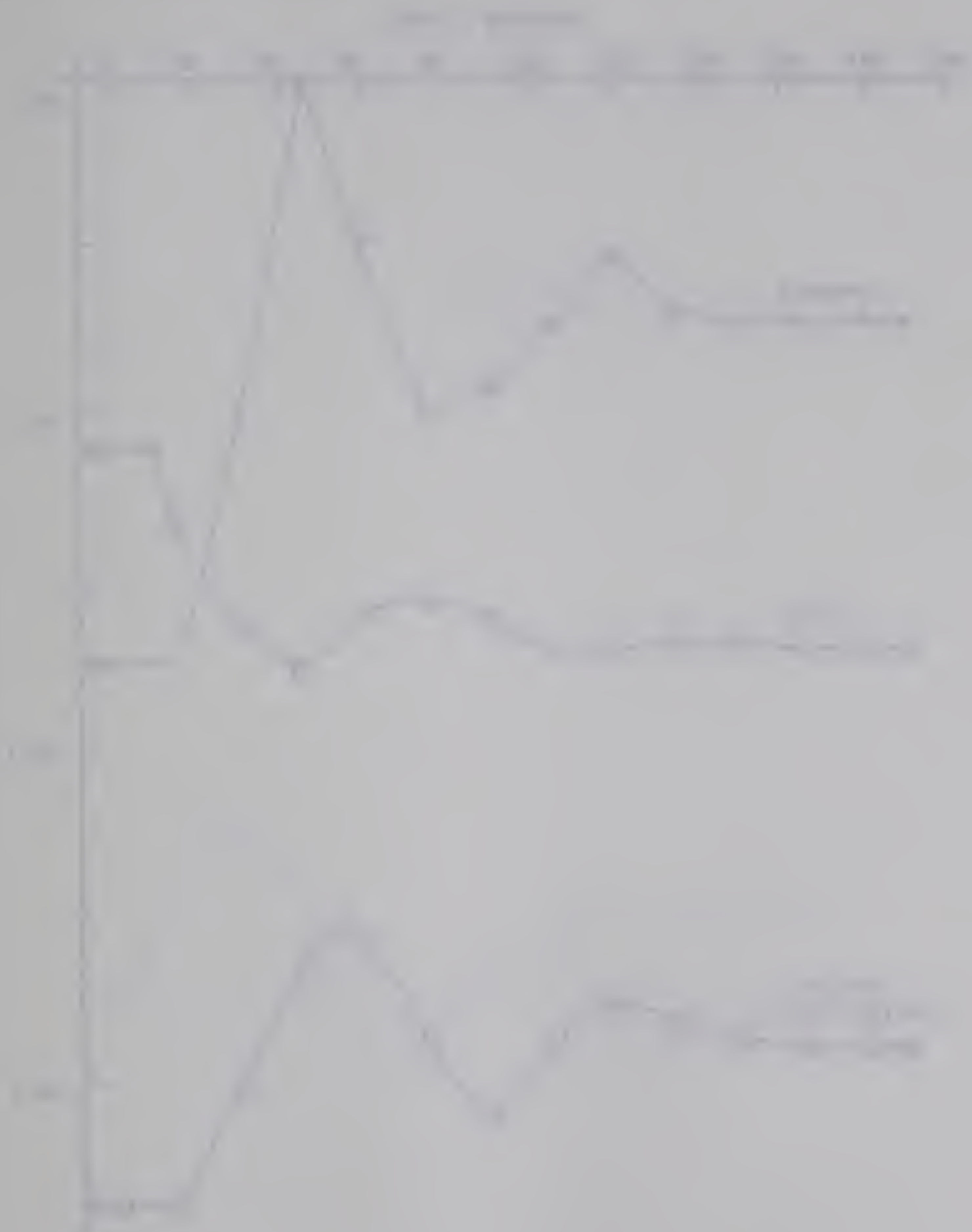


Figure 47 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN NO. 17FB

[illegible]



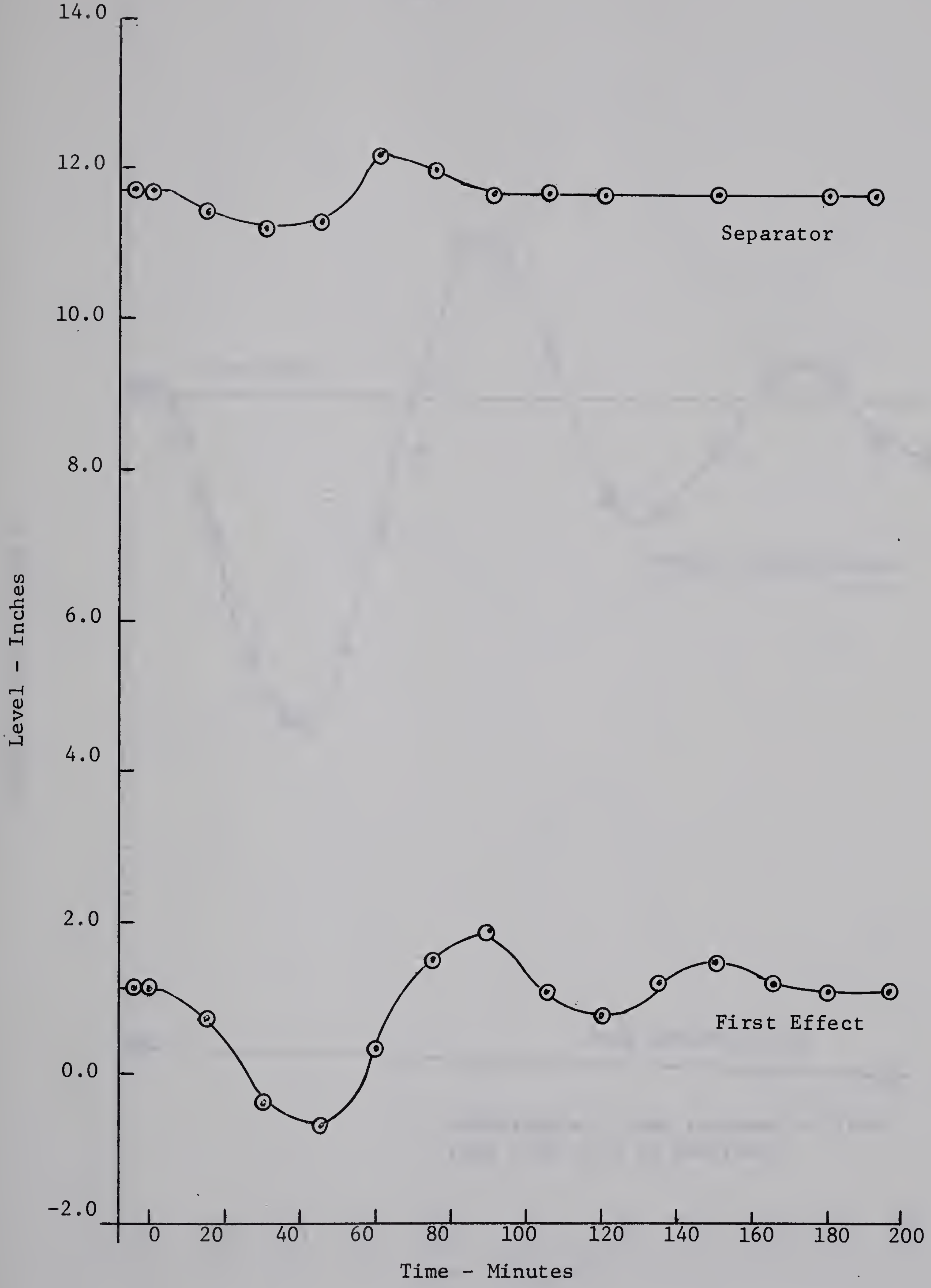
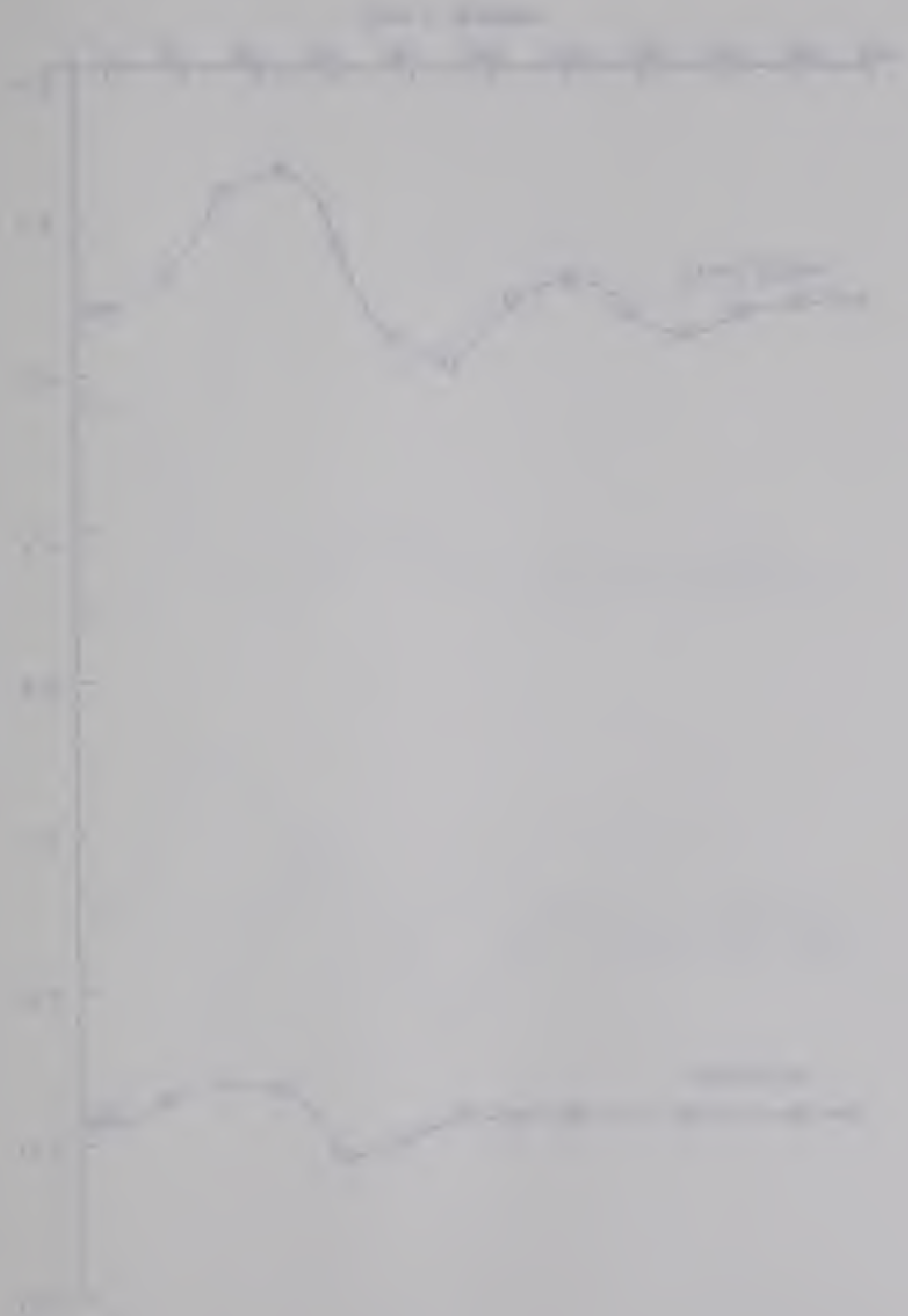


Figure 48 TRANSIENT RESPONSE OF SYSTEM LEVELS FOR RUN NO. 17FB

[illegible]



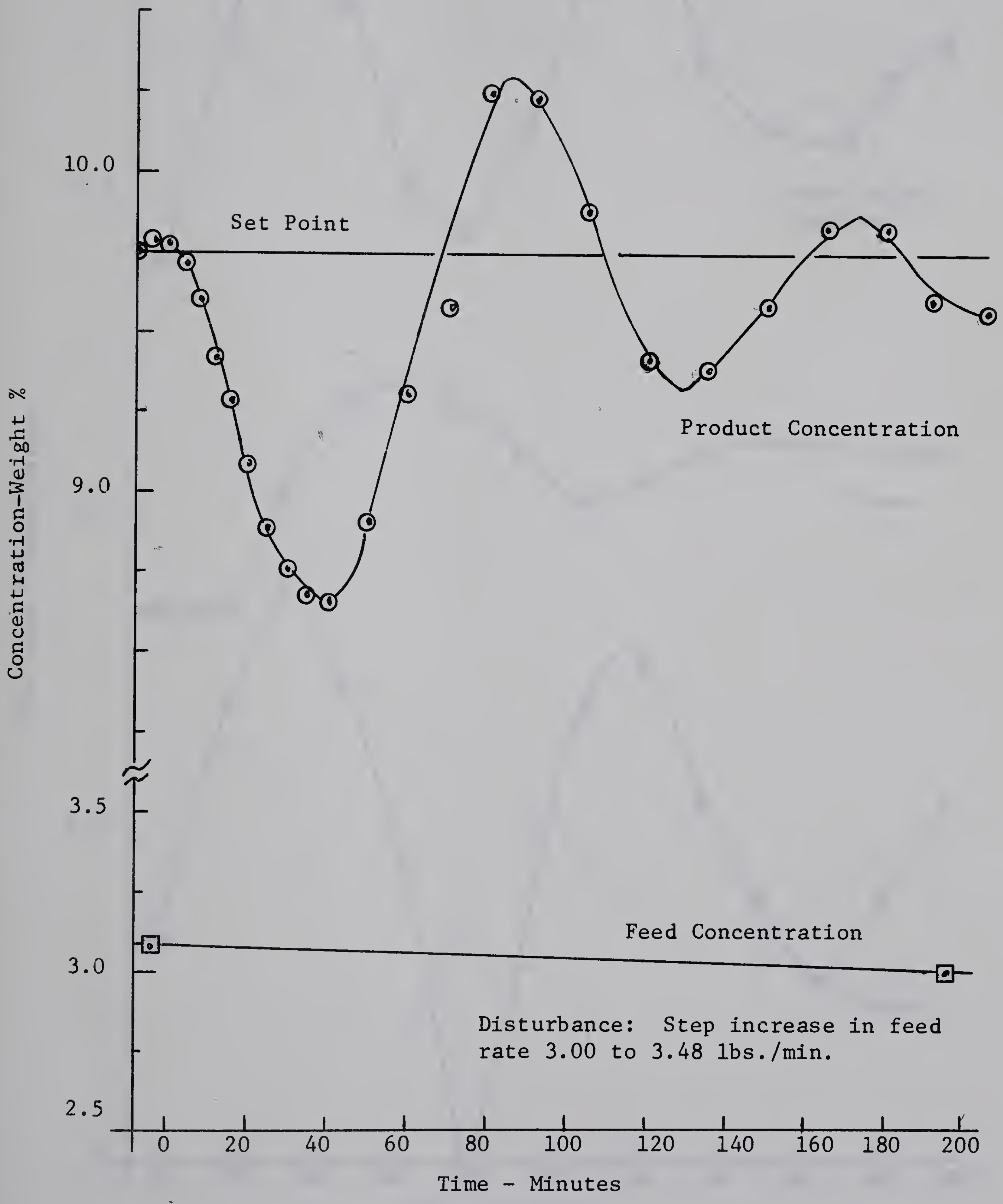
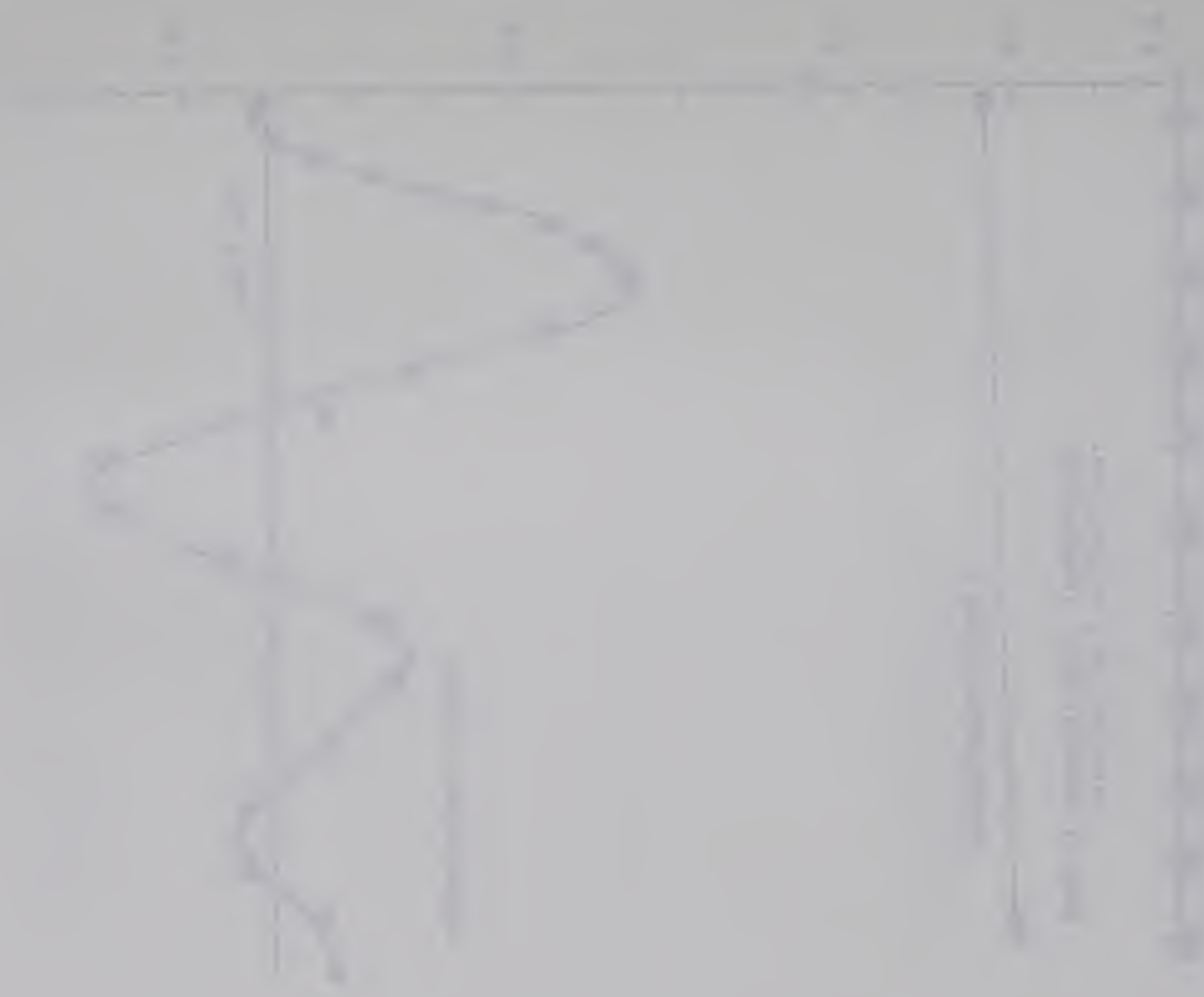


Figure 49 TRANSIENT CONCENTRATION RESPONSE FOR RUN NO. 18FB



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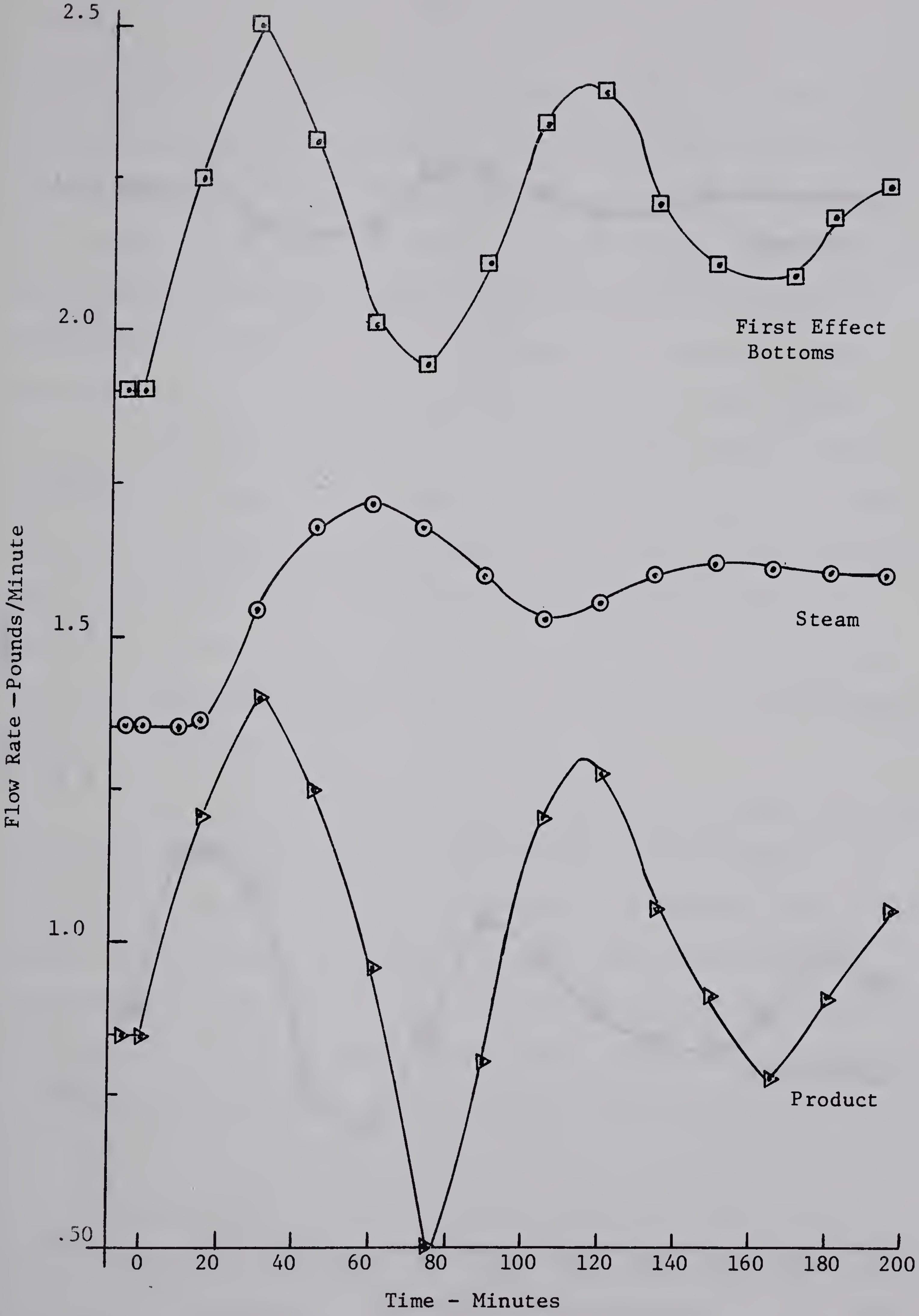


Figure 50 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN NO. 18FB

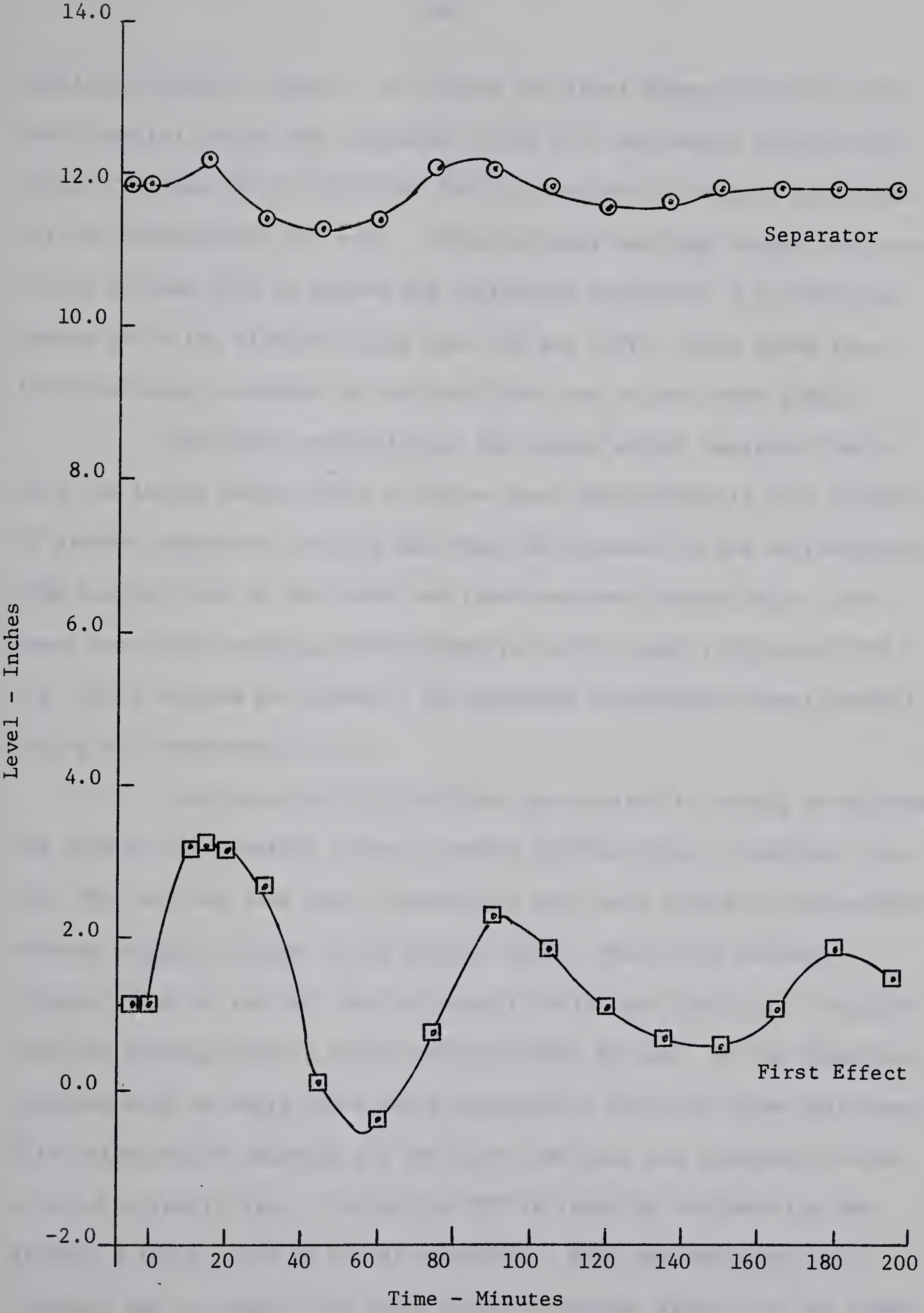
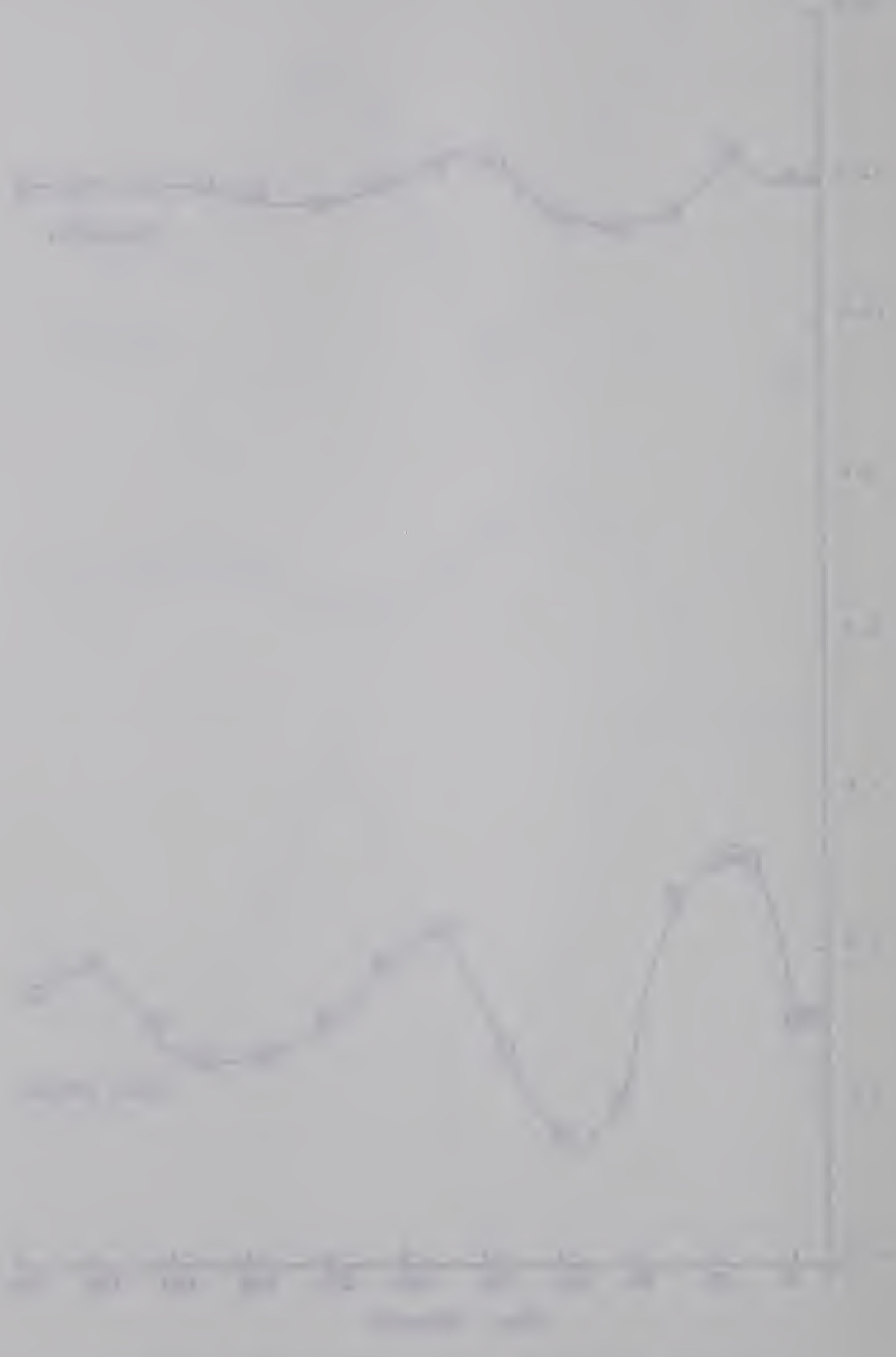


Figure 51 TRANSIENT RESPONSE OF SYSTEM LEVELS FOR RUN NO. 18FB



[illegible]

persisted slightly longer. To shorten the level transient period, the level control action was increased to 50% P.B. and twenty minutes per repeat for runs 10L to 11FFB and 50% P.B. and twelve minutes per repeat for the remainder of the runs. These narrower settings caused the first effect bottoms flow to exceed the calibrated maximum of 3.0 pounds per minute for a few minutes during runs 40L and 15FFB. Both these runs involved large increases in the feed flow rate to the first effect.

The level controller on the second effect separator had to hold the liquid level within a narrow range (approximately four inches) to prevent separator flooding and vapor entrainment in the recirculation pump suction line at the upper and lower extremes respectively. The level controller settings were therefore fairly tight (typically 120% P.B. and 5 minutes per repeat), but provided satisfactory level control during all experimental runs.

Considerable difficulty was encountered in trying to maintain the product flow control valve in smooth working order. Open loop runs 20L, 30L, 40L and some early feedforward runs were made with the original Foxboro Stabilflo valve in the product line. This valve had been cleaned prior to run 20L, but it began to stick and leak sugar solution from the packing after a relatively few hours of use. It was therefore replaced with an Annin valve which contained a built-in valve positioner. This valve worked smoothly for the first few runs but eventually began to stick slightly also. During run 9FF it stuck at one position and allowed a level build up in the separator. When the positioner air pressure was increased, the valve began to control again, but the sudden increase in flow as the valve opened to return the level to its set point

caused some distortion in the concentration response. Approximately one hour of the separator flow and level transients were omitted because of this error; the transients of Figure 32 and 33 were included only to indicate the initial and final steady state conditions for run 9FF. Prior to run 15FFB the air pressure had to be increased to this valve again. The flow disturbance that resulted, carried over into this run and resulted in a slight separator liquid level error at the start of this run. Fortunately, the product concentration was not noticeably affected by the slight adjustment to the product flow rate that followed the resetting of the positioner supply pressure. The product valve behaved favorably for the remainder of the experimental runs.

Aside from the problems of refractometer output drift and sticking valves, the evaporator equipment and instrumentation behaved in a fairly trouble-free manner. A small amount of oscillation or noise can be detected in the signals representing the first effect bottoms, product, and condenser flows, but this noise was considerably less than was reported by Andre [(2), pp. 109 and 4-14, 4-15]. Process records taken from run number 10FF are presented in Appendix E. It is evident that there is no particular difficulty in determining the values of the variables represented by these recordings. Where there was some oscillation in the recording, the chart reading was taken as the average of the extreme positions to which the pen travelled.

The cleaner chart records were mainly the result of three improvements to the evaporator system. The first was the addition of the level guages to the first effect and separator and the control of liquid

levels using level signals transmitted from these guages. They effectively damped out much of the noise in the first effect and separator. Thus both level records and flows manipulated by the two level controllers are not excessively noisy. All the pen dampers which had been installed by Andre for his experimental runs were removed following the installation of the level guages. The noise which did remain in the flow signals can be attributed to the tendency of the valves to stick once they become coated with the sugar solution. (When the Annin valve was first installed there was absolutely no oscillation in the product flow signal.) The valve positioners on the two valves operating in sugar solutions helped overcome their tendency to stick and kept the oscillation to a minimum amplitude.

The second main improvement was the installation of the high capacity centrifugal pump in the second effect recirculation line. Under normal operating conditions, there was no longer any cavitation in the recirculation pump or in the recirculation line, and the flow of fluid through the heating tubes was large and steady. It was felt that when the Moyno pump was recirculating liquid through the heating tubes, the flow of solution was not great enough to prevent slugging and surging of the solution through the tubes. These effects and occasional cavitation within the pump itself caused pressure disturbances in the second effect which appeared as noise in the first effect bottoms, the product, and the condenser condensate flow signals. This belief was supported by observations made during run number 10L. (See Appendix G). This run was made to determine the ability of the condenser condensate spillover level control system to function as the condenser pressure approached the

vacuum supply pressure. Condenser and second effect pressure was dropped approximately 50% to 3.0 psia: the condensate outflow increased to accommodate the increased boilup in the second effect and at no time did the condenser liquid level build up or threaten to flood. However, at the decreased operating pressure, cavitation resulted in the recirculation pump and the flow through the second effect heating tubes decreased and became rougher. The resulting surges in the recirculating flow resulted in noise peaks in the first effect bottoms, product, and condensate flows which were very similar to those observed on the typical process records in Appendix 4 of Andre's report.

The third improvement has already been referred to: it was the installation of spillover level control for the condenser discharge. There was some noise in the condensate flow signal following the switchover but certainly not enough to affect the ease of making reasonably accurate readings. Condenser flooding was no longer any problem once this condensate collection system was installed.

It is felt that a considerable improvement has been made to the general operation of the double effect evaporator as a result of this study. The development, construction, and integration of a relatively simple feedforward, product concentration controller with a conventional evaporator control system has also brought to light some of the problems which will no doubt be encountered when the evaporator is to be transferred to digital control.

Even though the Foxboro controllers on the evaporator and the feedforward concentration controller were not optimally tuned to

best cope with the evaporator process dynamics, it is hoped that the usefulness of at least some feedforward control for a process involving large system time constants has been demonstrated. All of the runs that were made using the feedforward system described in Chapter II and Appencix C have been included in this report. Some were unstable; others were marred by unfortunate minor process upsets, but all of them contribute towards an understanding of the problems and methods of obtaining a suitable control system for a process such as multiple effect evaporation.

The steady state model of the evaporating system which was programmed into the analog controller contained very little of the actual process dynamics. It therefore appears to be a very simple controller. The number of computing components required for a process model rises very quickly once the entire process dynamics and its many associated variables are brought into the program. Thus a model of the complexity of the one developed for this project is approaching the maximum manageable size for a control system built from individual components. Although this model succeeded in improving the ability of the evaporator control system to compensate for input disturbances in either feed flow or feed concentration, it is only a slight indication of the improvement to the control system which might result when the entire process dynamics are included in an accurate model of the evaporator.

The forthcoming installation of a digital process control computer will remove most of the model size limitations, and should allow a totally new approach for the automatic control of the double effect evaporator. A first logical step to make, might be to adapt the

evaporator model developed by Andre to a feedforward control system which could be integrated with the conventional evaporator controllers. Supervisory control, direct digital control, and then the development of an optimum evaporator control system are all projects which become realizable once the computing and memory capacity of a digital computer are added to the control system.

CHAPTER VIII

CONCLUSIONS

A feedforward product concentration controller has been designed for a general evaporating process, and interfaced with the electronic instrumentation on the unit operations laboratory double effect evaporator. The model used by the feedforward controller is based primarily on a steady state material balance around the evaporating unit. Feed flow rate, feed concentration and the product concentration set point are the process variables which the controller model uses to compute the steady state steam flow requirements of the unit. The feed concentration measurement signal enters the controller through a ten minute time constant circuit. This circuit simulates the concentration response of the first effect, which is essentially a first order system with a volume-to-throughput ratio of approximately ten minutes.

When product concentration control was completely supplied by the feedforward controller, the oscillatory period and the magnitude of concentration oscillations, following disturbances to feed rate or concentration, were decreased from those reported by Andre and observed during this study for similar runs utilizing feedback concentration control only. The concentration response following a product concentration set point change is essentially open loop under feedforward control. It is therefore slower than the response that can be obtained using a properly tuned three-mode feedback concentration controller.

The physical limitations of the evaporator model permitted small permanent offsets in the product concentration during some

experimental runs. The addition of negative feedback trimming to the feedforward controller eliminated these errors, and had the added advantage of decreasing the oscillatory period following disturbances to the feed rate or concentration. However, the combined feedforward-feedback control system could not provide suitable concentration control during product concentration set point changes. The resulting concentration response was stable but very slow in reaching the new set point.

This project has illustrated the suitability of applying feedforward control principles to evaporator control systems. Where the process model approximately described the dynamic steam requirements during transient periods, in addition to the steady state requirements, the product concentration response was very satisfactory. It is felt that further trial-and-error adjustment of the controller parameters could have further improved the concentration response following feed upsets. Such a program would have been very time consuming and could not have added any further conclusions to this study. Therefore no attempt was made to optimize the concentration controller settings.

A comparison of the concentration responses following product set point changes to the responses which followed feed upsets illustrates the desirability of including at least the major process dynamics in the feedforward model. When the computational power of a digital process computer becomes available later this year, the feedforward system might logically be extended to include the dynamic evaporator model developed by Andre (2). When this can be done, complete and accurate dynamic compensation for an upset in any process input will become a possibility.

Some significant improvements to the evaporator equipment and performance have also been achieved during this study.

1. The new condensate collection system added to the evaporator prevented condenser flooding under any operating conditions, and also allowed separate collection of all the condensate streams from the evaporator for volumetric calibration.

2. A new recirculation system was installed in the forced circulation evaporator. The increase in recirculation flow rate that resulted produced a substantial increase in the heat transfer coefficient at the second effect.

3. The block and drain valves installed below some of the process flow transmitters allowed calibration runs and on-line checks to be carried out quickly and conveniently.

4. The external level guages added to the first effect and separator level control systems dampened out much of the noise in these level signals. This addition, and the effect of a smoother recirculation flow in the second effect, allowed the removal of mechanical pen dampers from the process recorders. All process records were relatively noise free during this study.

5. A steam condensate trap added to the inlet steam line at the first effect part way through the experimental program, effectively removed condensate from the supply steam and thereby improved the accuracy of the heat balance calculations.

6. The removal of the condenser vacuum control valve from the vacuum supply line and its relocation in the atmospheric bleed to the

vacuum provided excellent pressure control and removed the necessity of manually adjusting the atmospheric bleed valve.

A. Future Work

It is evident that there is much more to be learned about multiple effect evaporators and their control systems. Once the evaporator is operating under direct digital control it will become much easier to widen the scope of this control system study. Eventually, after the evaporator has been accurately modelled and many combinations of feed-forward and feedback control systems have been investigated, an adaptive control system may be developed which will optimally control the evaporating unit under any operating conditions.

In view of the fact that the evaporator is to be dismantled and rebuilt later this year in a new Chemical Engineering dynamics research laboratory, a number of recommendations regarding evaporator operation are of immediate interest. These recommendations are essentially by-products of the completed control system study and, if adopted, should simplify the evaporator operation. They are:

1. A continuous feed system should be designed to replace the present batch feed operation. This feed system should contain provisions for both temperature and concentration control of the feed stream. The concentration controller should be capable of supplying rapid changes to the feed concentration; the temperature controller should be able to hold the feed temperature within a very narrow temperature range so that the feed refractometer can be used without temperature compensation if so desired.

2. Hand valves should be installed in process inlet and output lines (where necessary) so that the evaporator can be completely isolated from process tankage during pressure checks on the unit. It is also important that air leaks be kept to a minimum in the first effect since the air accumulates in the steam chest of the second effect, where it can only be removed by purging some of the steam from the steam chest.

3. The need for dependable concentration measurements in both the feed and product steams was great during this study, and will be even more important if a continuous feed system is to be developed. If the refractometers cannot be depended upon to hold their calibration, they will have to be replaced with concentration measuring instruments that will.

4. It is often advantageous to be able to make an overall enthalpy balance around a process unit. If an independent overall enthalpy balance is to be made on the evaporator, it will be necessary to install a flow transmitter or orifice run and a number of thermocouples in the cooling water line.

5. Stainless steel pipe is not essential for water and sugar-water duty and could be replaced by cheaper copper tubing when the evaporator is rebuilt. Plastic piping has proven to be a headache and any presently remaining on the evaporator should also be replaced with copper tubing.

6. The Fostoria Dynapump inlet lines should be enlarged to a minimum of one inch tubing to allow proper suction for these pumps.

7. The possibility of using some other solute than sugar should be considered as this might be the easiest method of avoiding the

problems which the sticky sugar solutions have caused. Any solute substituted should be non-volatile to prevent the loss of solute in the overhead vapor streams from each effect.

8. The in-line flow transmitter and dp cell units metering the hot solution flows from the first and second effects should be replaced with orifice runs and separate dp cells. The dp cells could then operate at ambient temperature.

If the above recommendations are successfully carried out the evaporator operation will become essentially automatic. It should then be possible to obtain more useful experimental data from the evaporator during any period of operation than is now possible. These changes, when combined with those already completed on the double effect evaporator, should make it much easier for the operator to run the evaporator instead of vice versa.

NOMENCLATURE

A. General

W	Mass flow rate (lbs./min.)
C	Sugar concentration (weight percent sugar)
S	Steam flow rate (lbs./min.)
O	Overhead vapor flow (lbs./min.)
T	Temperature (°F)
Λ	Heat of vaporization (Btu/lb.)
HL	Heat losses (Btu/min.)
H	Heat supply per pound of steam condensed (Btu/lb.)

B. Symbols Pertinent to Wiring Diagrams

Symbol	Name	Voltage Range
V_{co}	Voltage representing feed concentration	(2-10 volts)
V_{cl}	Voltage representing concentration in first effect	(2-10 volts)
V_{cpsp}	Voltage representing product concentration set point	(1-5 volts)
V_{wo}	Voltage representing feed flow rate	(2-10 volts)
V_s	Voltage representing steam flow set point	(1-5 volts)
C_o	Voltage equivalent of feed concentration	(2-6 volts)
C_{psp}	Voltage equivalent of product concentration set point	(4-12 volts)
W_o	Voltage equivalent of feed flow	(1.16 - 5.38 volts)
S	Voltage equivalent of required steam flow	(0 - 2.88 volts)

Symbol	Name	Voltage Range
K_c	Controller gain	
T	Time constant of first effect	(10 minutes)

C. Potentiometer Settings

K_s	Ratio of steam required per pound of water evaporated	
L	Latent heat term	
B	Steam set point bias	

D. Subscripts

o	Refers to feed stream
1	Refers to the first effect
2	Refers to the second effect
12	Refers to inter-stage connections
p	Refers to the concentrated sugar product stream
sp	Refers to a set point
s	Refers to steam flow
MB	Denotes material balance
LH	Denotes latent heat
HL	Denotes heat loss

A Bar "—" over a symbol implies that it is an average value.

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APPENDIX A

PROCESS FLOW SHEET AND EVAPORATOR STARTUP PROCEDURE

The basic evaporation process equipment and its associated Foxboro instrumentation have undergone a number of significant changes since Andre described the double effect evaporator. Therefore, it was felt that a complete process flow sheet, including all evaporator auxilliary equipment, would be valuable since the evaporator is to be completely dismantled and then rebuilt in a new laboratory. Figure A-1 is a complete flow diagram of the evaporating system as it was during the period when the experimental runs reported in this thesis were made. The listing of automatic and manual control valves which follows the flow sheet is included to aid in their identification and location on the flow sheet.

A. Process Flow Sheet

1. Symbols Used

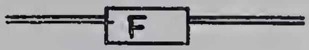
(a) System Inputs

⊙———	Steam Supply
▶———	Cooling Water Supply
⬡———	Vacuum Supply
●═══	Feed Supply
⊡———	Air Supply

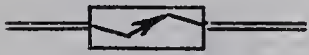
(b) System Outputs

═══▶	Condensate to Tankage
═══▷	Product to Tankage
↓Y	Water to Sewer

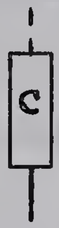
(c) Special Equipment



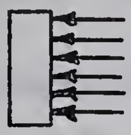
Filter



Heat Exchanger

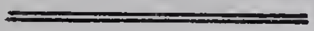


Calibration Volume

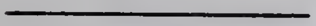


Vacuum Manifold for Sampling Devices (First Effect)

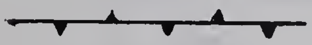
(d) Process Lines



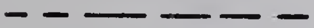
Main Process Lines: Vapor, Condensate and Solution



Secondary Process Lines



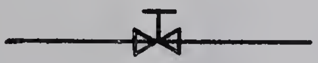
Steam Line



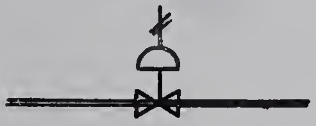
Vacuum Line

(e) Instrumentation and Control Equipment

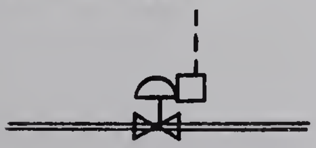
(i) Valves



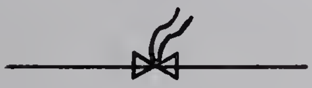
Hand Valve



Pneumatic Control Valve

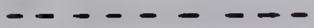


Pneumatic Control Valve with Valve Positioner

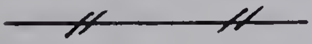


Electric Solenoid Valve

(ii) Control Signal Lines

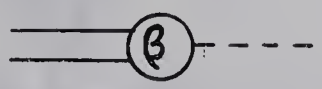


Electric Signal, Voltage or Current

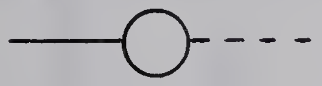


Pneumatic Signal

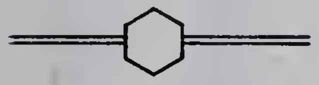
(iii) Transmitters



Electric Differential Pressure Transmitter
(a) β = LT Level Transmitter
(b) β = PT Pressure Transmitter
(c) β = DVPT Vapor Pressure Transmitter
(d) β = FT Flow Transmitter

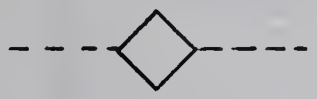


Temperature Transmitter (Thermocouple)

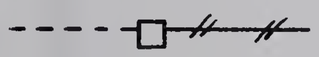


Refractometer

(iv) Transducers

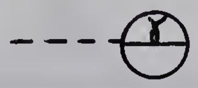


Voltage to Current

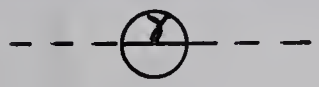


Current to Air

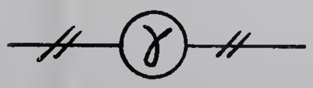
(v) Recorders (R), Controllers (C),* Recording Controllers (RC),* and Indicaters (I)



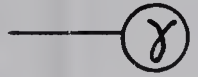
Pannel Mounted Recorder



Pannel Mounted Controller or Recording Controller



Local Controller or Indicating Controller (Pneumatic)



Local Indicator

The symbol γ may represent level (L), flow (F), Composition (C), Temperature (T), or Pressure (P).

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* The valves associated with a particular control loop will be identified by the number of that loop with a "V" affixed to the controller function. eg. The valve manipulated by FRC-3 (Feed Flow) is denoted FRCV-3.

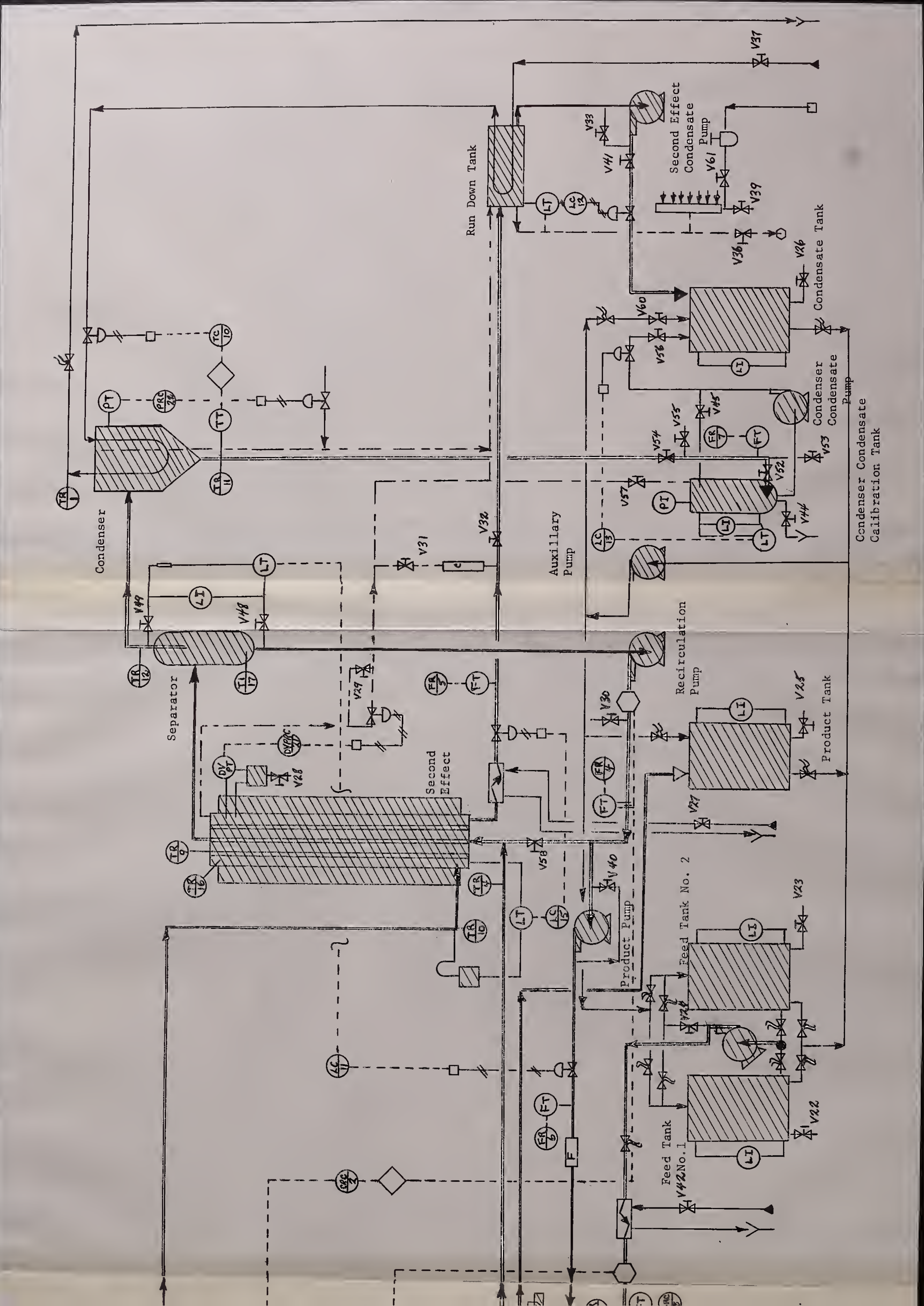

Condenser
Separator
Second Effect
Run Down Tank
Second Effect Condensate Pump
Condensate Tank
Condenser Condensate Pump
Condenser Condensate Calibration Tank
Auxillary Pump
Recirculation Pump
Product Tank
Product Pump
Feed Tank No. 2
Feed Tank No.1
TR 1
TR 12
TI 17
TR 9
TR 16
TR 10
TR 11
PT
PRC 24
TT
TC 10
LT
LI
LC 12
LC 13
LC 15
LC 11
FR 5
FR 4
FR 6
FR 7
FT
F
DV PT
DVRC
V22
V23
V25
V26
V27
V28
V29
V30
V31
V32
V33
V36
V37
V39
V40
V41
V42
V44
V45
V48
V49
V52
V53
V54
V55
V56
V57
V58
V60
V61

Table A-1

AUTOMATIC CONTROL VALVES

Valve Number	Control Variable	Manipulated Variable	Controller Actions*	Comments
FRCV-1	Steam Flow	Steam Flow (Pressure)	P-I	
FRCV-2	1st Effect Bottoms Flow	1st Effect Bottoms Flow	P-I	1st effect level controller cascaded to set point of FRC-2. Valve positioner on valve.
FRCV-3	Feed Flow	Feed Flow	P-I	Orifice size .0995"
LCV-10	Steam Chest Condensate Level	Condensate Flow	P-I	Valve positioner on valve.
LRCV-11	Separator Liquid Level	Product Flow Rate	P-I	External level guage Supplies level signal
LCV-12	Run Down Tank Liquid Level	Condensate Flow to Tankage		Pneumatic loop
LCV-13	Condensate Calibration Tank Level	Condensate Flow to Tankage	P-I	
LCV-15	2nd Effect Condensate Level	2nd Effect Condensate Flow	P-I	
TCV-10	Condenser Condensate Temperature	Cooling Water Flow	P-I	Used on manual control only
DVPRCV-21	Inerts Content of 2nd Effect Steam	Vacuum Purge	P-I	
PRCV-22	Condenser Pressure	Air Bleed to Vacuum	P-I	

*"P" Denotes Proportional Action
"I" Denotes Integral Action
"D" Denotes Derivative Action

Table A-2

AUTOMATIC RECORDERS AND CONTROLLERS

Valve Number	Control Variable	Manipulated Variable	Controller Action	Comments
FR-4	2nd Effect Circulating Flow			Square edge orifice diameter = 0.875". Pressure Differential = 200 inches of water
CR-1	Feed Concentration			
CRC-2	Product Concentration	FRC-1 Set Point	P-I-D	Cascade control of FRC-1 used to provide feedback product concentration control
FR-5	2nd Effect Condensate Flow			Orifice = 0.0349"
FR-6	2nd Effect Bottoms Product Flow			Orifice = 0.0995"
FR-7	Condenser Condensate Flow			Orifice = 0.114" Pressure differential = 20 inches of water
PR-20	1st Effect Pressure			
LRC-14	Liquid Level in 1st Effect	FRC-2 Set Point	P-I	Level transmitter receives signal from an external level guage

Table A-3

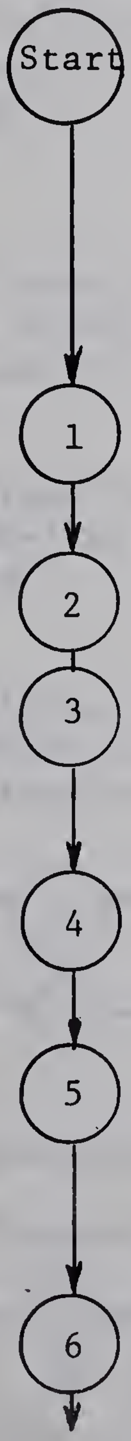
MANUALLY OPERATED VALVES

V20	Steam inlet valve
V21	First effect pump recycle valve
V22	Feed tank No. 1 drain valve
V23	Feed tank No. 2 drain valve
V24	Feed pump recycle valve
V25	Product tank drain valve
V26	Condensate tank drain valve
V27	Second effect condensate cooler inlet water valve
V28	Second effect water chamber blowdown valve
V29	Second effect purge bypass valve
V30	Product sample valve
V31	Calibration vessel purge valve
V32	Second effect condensate block valve (calibration)
V33	Second effect condensate pump recycle valve
V34	First effect bottoms flow block valve
V35	First effect bottoms flow calibration valve
V36	Vacuum inlet valve
V37	Cooling water inlet valve
V38	Steam condensate calibration tank drain valve
V39	Vacuum manifold air intake
V40	Product pump recirculation valve
V41	Second effect condensate pump block valve
V42	Feed cooler water inlet valve
V43	Steam chest vent valve on first effect
V44	Condenser condensate calibration tank drain valve
V45	Condenser condensate tank recirculation valve
V46	First effect level guage inlet valve (bottom)
V47	First effect level guage inlet valve (top)
V48	Separator level guage inlet valve (bottom)
V49	Separator level guage inlet valve (top)
V50	Steam Condensate tank calibration valve
V51	Steam condensate tank inlet valve
V52, V54	Condenser condensate tank inlet valves
V53, V55	Condenser condensate tank calibration valves
V56	Condenser condensate tank outlet valve
V57	Condenser condensate tank vacuum valve
V58	Second effect recirculation line throttle valve
V59	First effect pump isolation valve
V60	Condensate tank inlet valve
V61	Evaporator pressure testing valve
V62	Condensate trap drain valve

B. Evaporator Start-up and Shutdown

1. Evaporator Start-up Procedure

The complex procedure of evaporator start-up can best be visualized by presenting the instructions in the form of a block diagram. Each step should be carried out as rapidly as is possible once the preceeding step has been completed. Some of the initial operations can be carried out simultaneously at the operators discretion. There must be liquid in the run down tank before the evaporator can be successfully put into operation.



Check Valve Positions: Open:
V32, V35, V40, V51, V52, V54,
V34, V45, V46, V47, V48, V49,
V57, V24, V43
Closed: V22, V23, V25, V26,
V31, V38, V44, V50, V53, V55,
V56.

Turn on Power to Controllers,
Refractometers, Solenoids and
Chart Drives

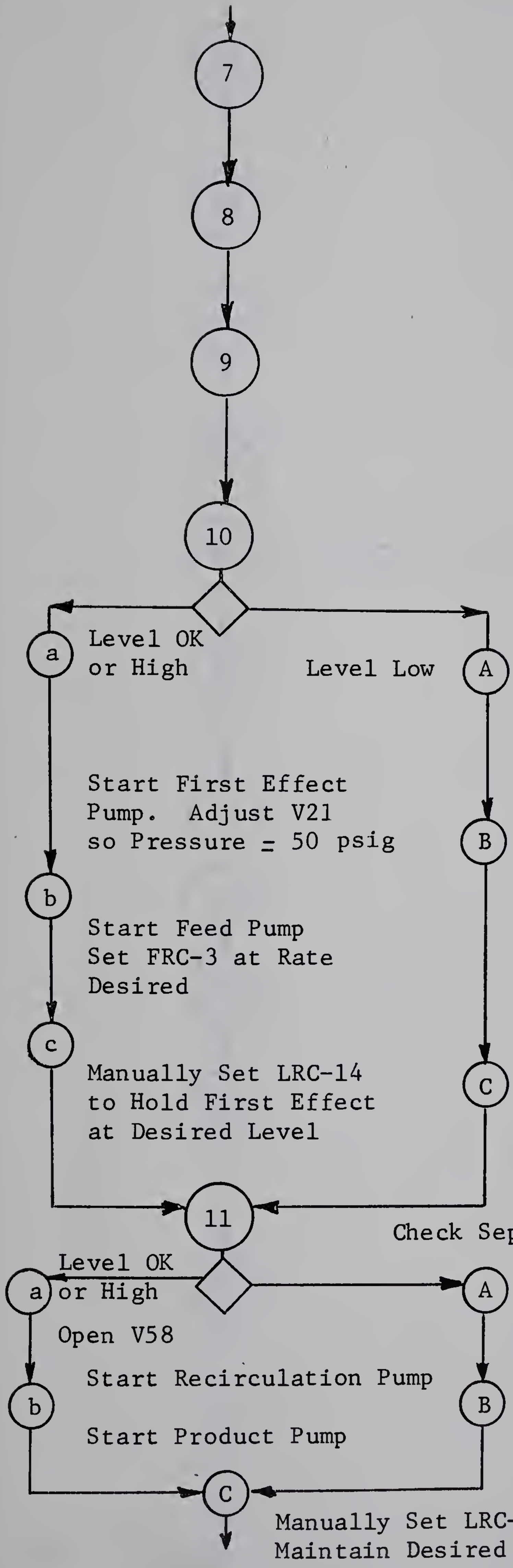
Turn on Air Solenoid

Open Steam Valve to Ejector.
Adjust Water Quench Valve to
1/4 Open

Set Knockout Drum Pressure
at 55 psig

Bleed Level Transmitters
(must have liquid in first
effect, run down tank and
separator)

Put on Manual Control and
Close: LRC-14, LRC-11, LC-15



7 Turn on Vacuum Solenoid
Open V36, V29
Close V39, V28

8 Set PRC-22 on Automatic Control
Set Point = 20%

9 Turn on Second Effect Condensate Pump. Adjust V33 and/or V41 to put Outlet Pressure to 35 psig

10 Check First Effect Level

Level OK or High

a Start First Effect Pump. Adjust V21 so Pressure = 50 psig

b Start Feed Pump
Set FRC-3 at Rate Desired

c Manually Set LRC-14 to Hold First Effect at Desired Level

Level Low

A Set FRC-3 at 100
Start Feed Pump

B Start First Effect Pump when Level Begins to Build up. Adjust V21 so Pressure = 50 psig

C Lower Feed Flow to Desired Rate
Manually set LRC-14 to Hold First Effect at Desired Level

11 Check Separator Level

Level OK or High

a Open V58

b Start Recirculation Pump
Start Product Pump

A Leave LRC-11 Closed Until Level Builds Up

B Open V58
Start Recirculation Pump
Start Product Pump

C Manually Set LRC-11 to Maintain Desired Separator Level

12 Set DVPRC-21 on Automatic
Control
Set Point = 15%

13 Turn on Water
Solenoid

14 Set TC-10 on Manual
Control with Valve
> 50% open

15 Open V42

16 Open V20, Close V43
After Steam Escapes

17 Set FRC-1 on Automatic Control
Set Point = 50% (Lower to 40%
in 90 min.)

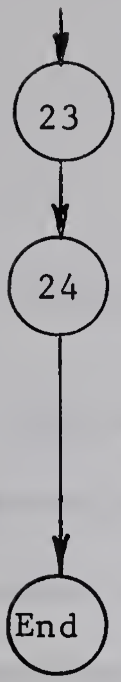
18 Clean Refractometer Lenses

19 Open V27 (1/2 Turn)

20 As Purge Bypass Line
Heats up, Close V29

21 As Condensate Builds Up in
Second Effect, Manually
Adjust LC-15 to Maintain a
Low Condensate Level

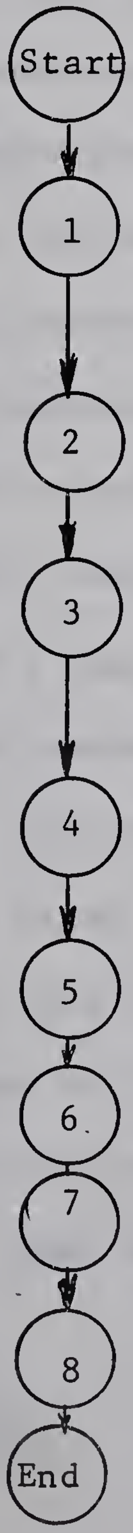
22 Check Liquid Levels: Adjust
Manual Settings if Necessary.
When Levels Stabilize, Balance
Controller and Switch to Automatic
Level Control.
Set Points:
LRC-14 = 25%
LC-15 = 5%
LRC-11 = 25%



Switch on Temperature
Recorder

Optional
After Product Concentration
Stabilizes Balance CRC-2 or
Feedforward Computer and
Cascade Composition Control
Signal to FRC-1

2. Evaporator Shut-Down Procedure



Switch PRC-22 to Manual
and Open

Turn off Vacuum Solenoid
Open V39, V29, V28
Close V36

Close V20
Open V43

Shut off Feed, First Effect,
Second Effect Condensate,
Circulation, and Product Pumps

Turn off Water Solenoid
Close V27, V42

Switch to Manual Control and Close:
LRC-14, LC-15, LRC-11

Turn off Air Solenoid

Turn off Power to Solenoids and
Chart Drives (Power to Refractometers
and Controllers may be left on)

Shut off Steam and Water to Steam
Ejector

APPENDIX B

CALIBRATION OF EVAPORATOR INSTRUMENTATION

The calibration data presented in Figures B-1 through B-15 and in Table B-2 are largely self-explanatory. Steam and feed flow and feed and product concentration calibration curves had to be programmed into the feedforward composition controller. A function generator consisting of several function segments and an add-subtract module would have been required to reproduce the non-linearities in each curve. The number of computing modules available was just sufficient to generate a linear approximation of each curve. The linear curves can only be used over a limited range to represent the parabolic flow calibration curves, but represent both concentration calibration curves very well. The linear approximations are presented in Figures B-3, B-5, B-14, B-12, and B-13 for steam flow, feed flow, feed concentration and product concentration respectively. Table B-1 includes the equations of these curves, their suggested ranges for application, and the maximum error which may be encountered in those ranges.

Two separate calibration curves are presented in Figures B-1 and B-2 for steam flow and in Figure B-4 for feed flow. The original calibration runs and open loop runs 20L, 30L, and 40L were made with the zero settings on the feed and steam flow transmitters slightly in error. Following these runs, all flow transmitters were checked and rezeroed, reset and then recalibrated if necessary.

Table B-1

LINEAR REPRESENTATION OF CALIBRATION CURVES

Variable	See Figure	Voltage Representation	Physical Range	Conversion Equation	Recommended Range	Maximum Error
Feed Flow	B-5	V_{wo}: 2-10 v.	W_o: 0.0-5.3 lb./min.	$W_o = 0.5375\ V_{wo} + 0.655$	V_{wo} = 4-7 v.	1.8% at $V_{wo} = 4.0$
Feed Concentration	B-14	V_{co}: 2-10 v.	C_o: 1.95 - 5.90%	$C_o = 0.4875\ V_{co} + 0.975$	V_{co} = 2-10 v.	0.0%
Product Concentration[a]	B-12	V_{cp}: 1-5 v.	C_p: 4.0 - 12.0%	$C_p = 2.0\ V_{cp} + 2.0$	V_{cp} = 1 - 4.8 v.	2.5% at V_{cp} = 4.75 v.
Steam Flow Set Point	B-3	V_s: 1-5 v.	S: 0.0 - 2.0 lb./min.	$V_s = 3.226S - 1.774$	V_s = 1.64 - 4.0 v.	5.0% at V_s = 1.64 v.

---

[a] Product Composition Set Point is represented by the same voltage range (V_{cpsp} = 1 to 5 volts) and therefore the same conversion equation applies.

The new calibration curve for steam flow was significantly lower than the original curve, as Figure B-3 clearly indicates. Unfortunately, the new calibration curve was not programmed into the feedforward concentration controller at the time the transmitter zero was reset. Steam flows were therefore below the flow rate called for by the controller. However, the balancing procedure outlined in Appendix C automatically eliminated the steam flow error before each run. Since the calibration error is essentially a negative bias, the bias term set in the feedforward controller during balancing would be increased by an amount just sufficient to cancel out the calibration error.

The feed flow calibration curve illustrated in Figure B-5 was programmed into the controller after the transmitter zero was reset. This curve represents the actual feed flow very accurately within the range from 30% to 50% chart. All experimental runs were made using feed rates within this range. Figures B-12 and B-14 indicate that the concentration calibration curves programmed into the feedforward controller represent the actual concentrations very well.

The four calibration curves discussed above are the only ones required by the feedforward concentration controller. Since they are acceptably accurate, or at worst, corrected automatically for inherent errors, it is believed that the feedforward controller output did not at any time contain large errors which could be directly attributed to incorrect conversion equations.

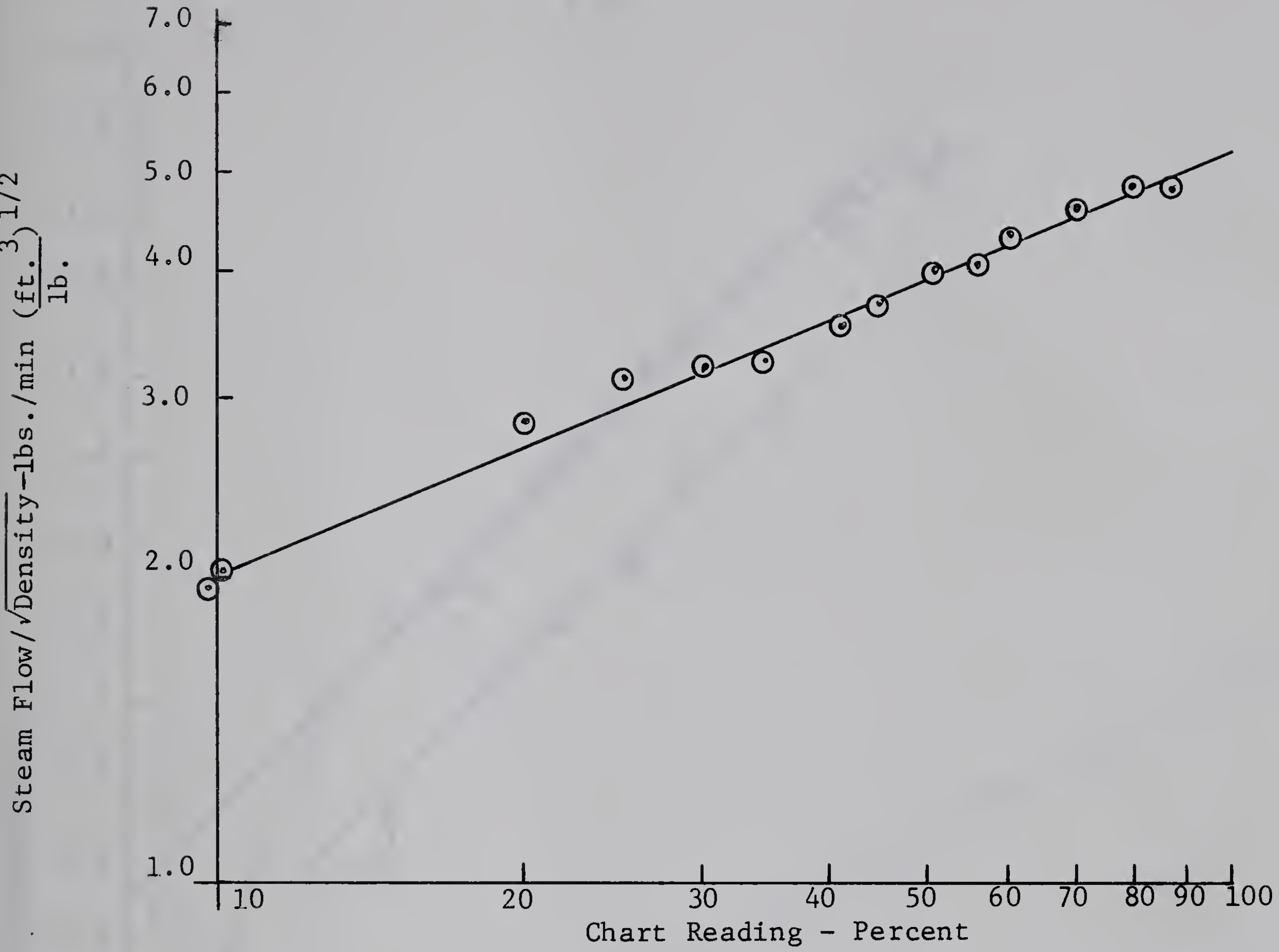


Figure B-1 CALIBRATION: STEAM FLOW TO FIRST EFFECT FOR RUNS 20L, 30L & 40L

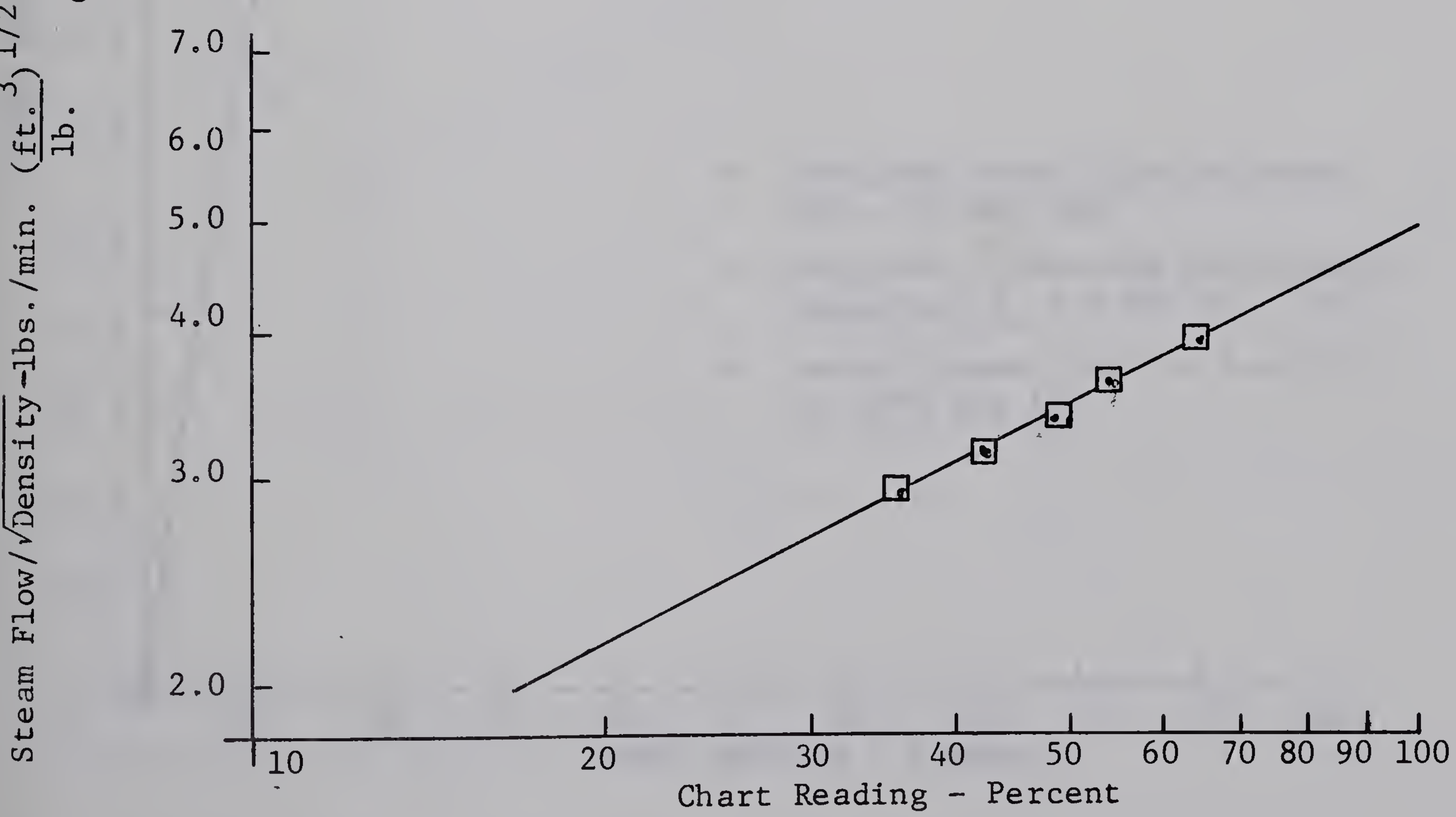


Figure B-2 CALIBRATION: STEAM FLOW TO FIRST EFFECT FOR RUNS 10L & 5FF to 18FB

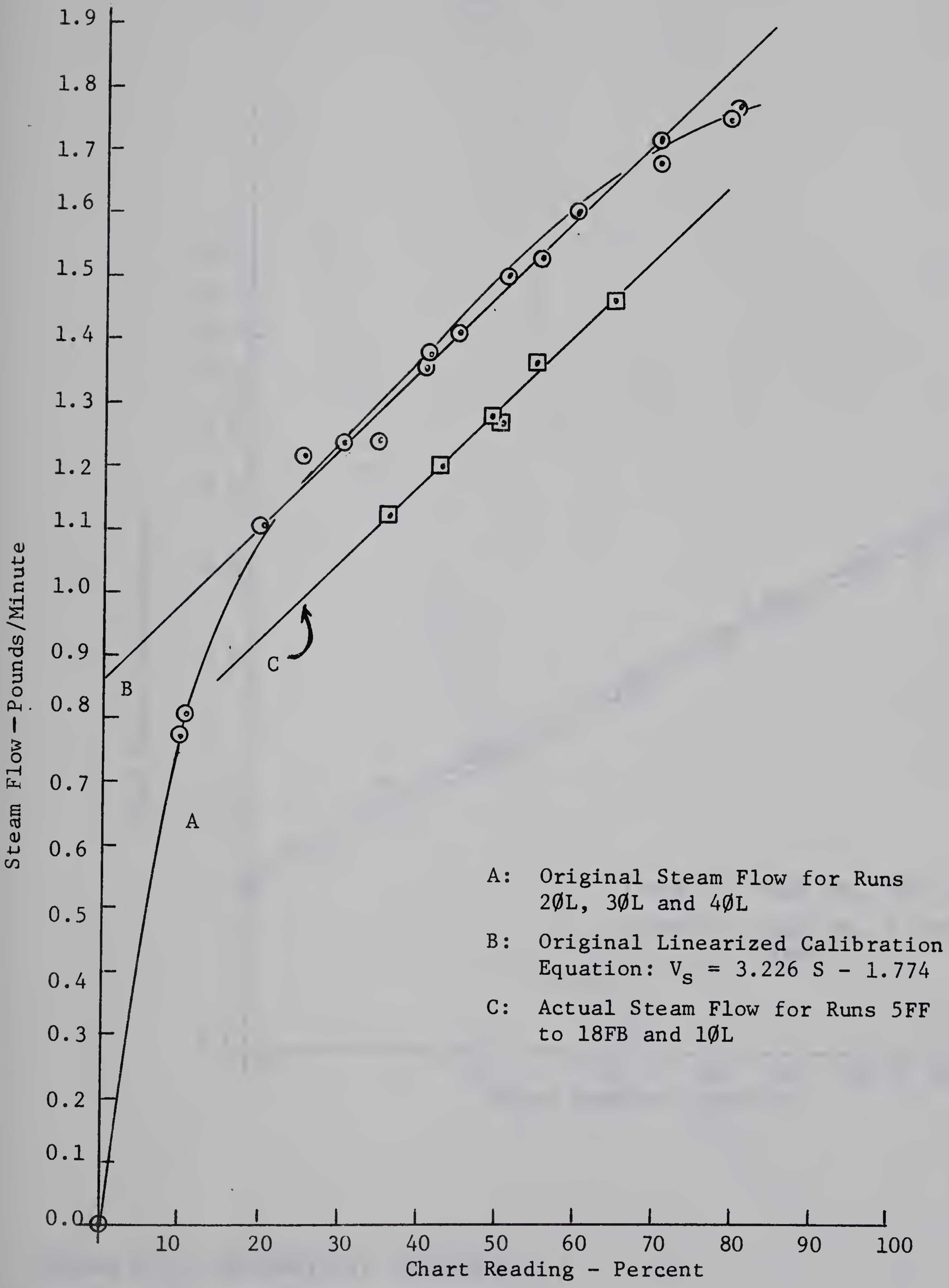


Figure B-3 CALIBRATION: STEAM FLOW TO FIRST EFFECT

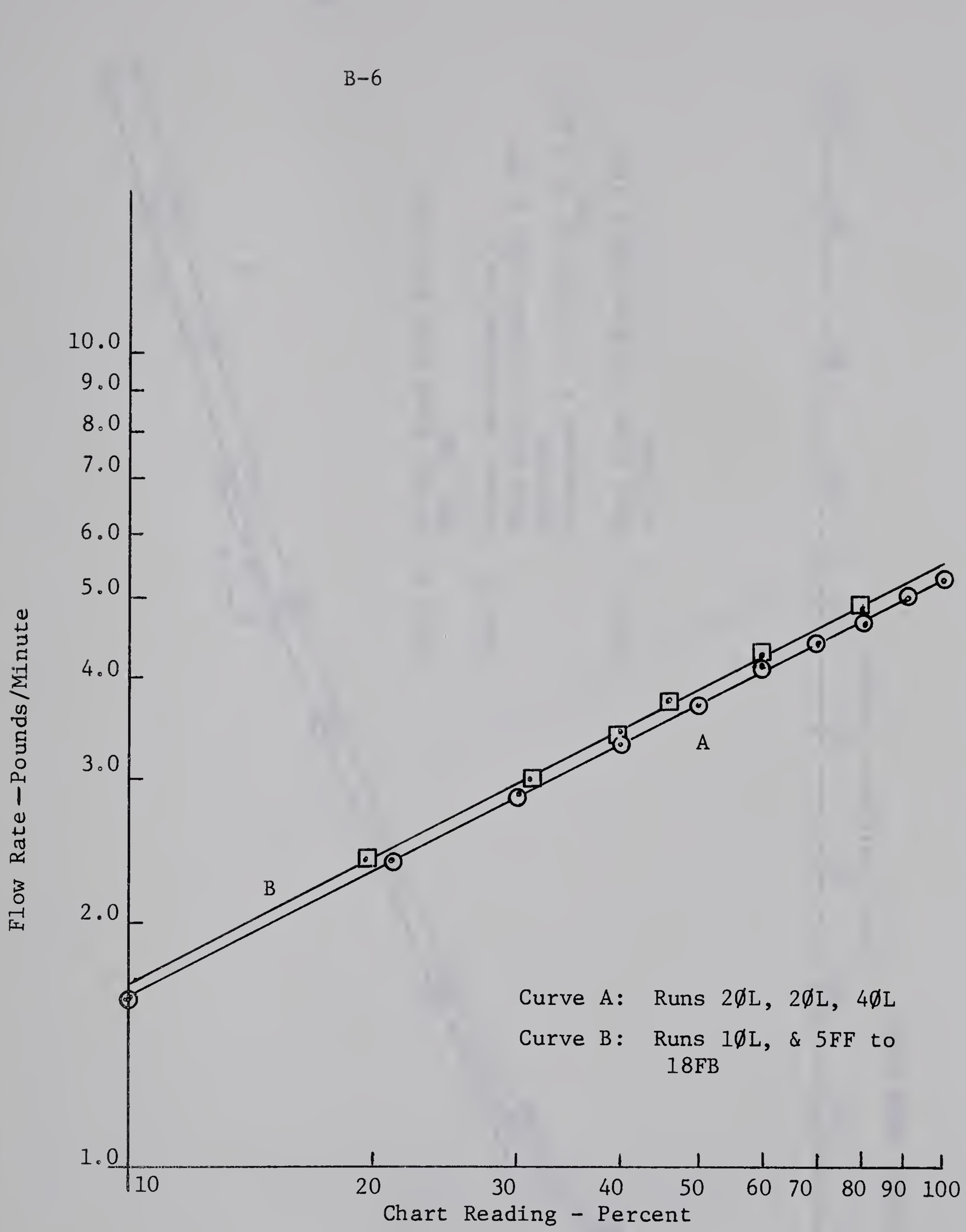


Figure B-4 CALIBRATION: FEED FLOW

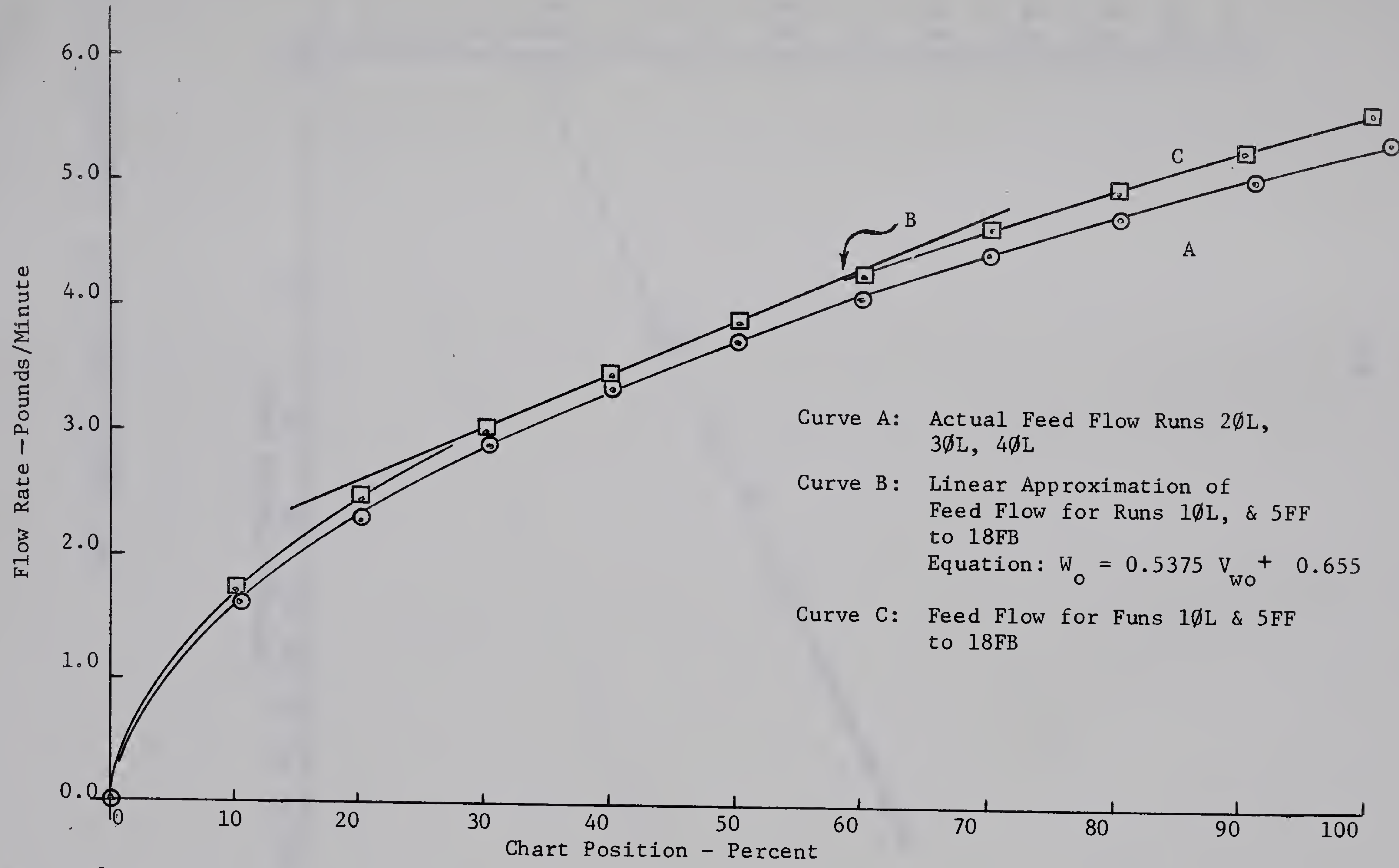


Figure B-5 CALIBRATION: FEED FLOW

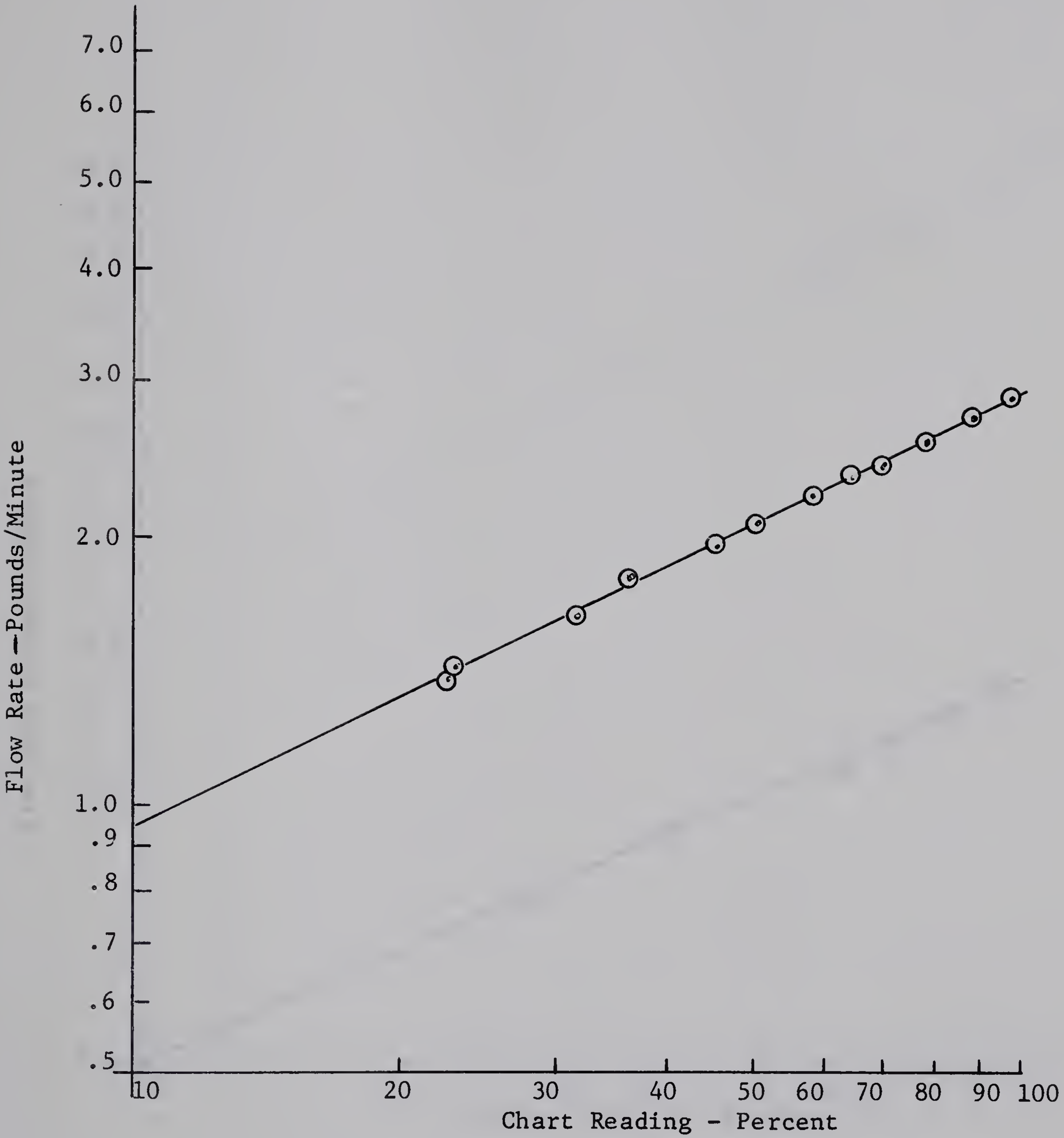


Figure B-6 CALIBRATION: FIRST EFFECT BOTTOMS FLOW

[illegible]

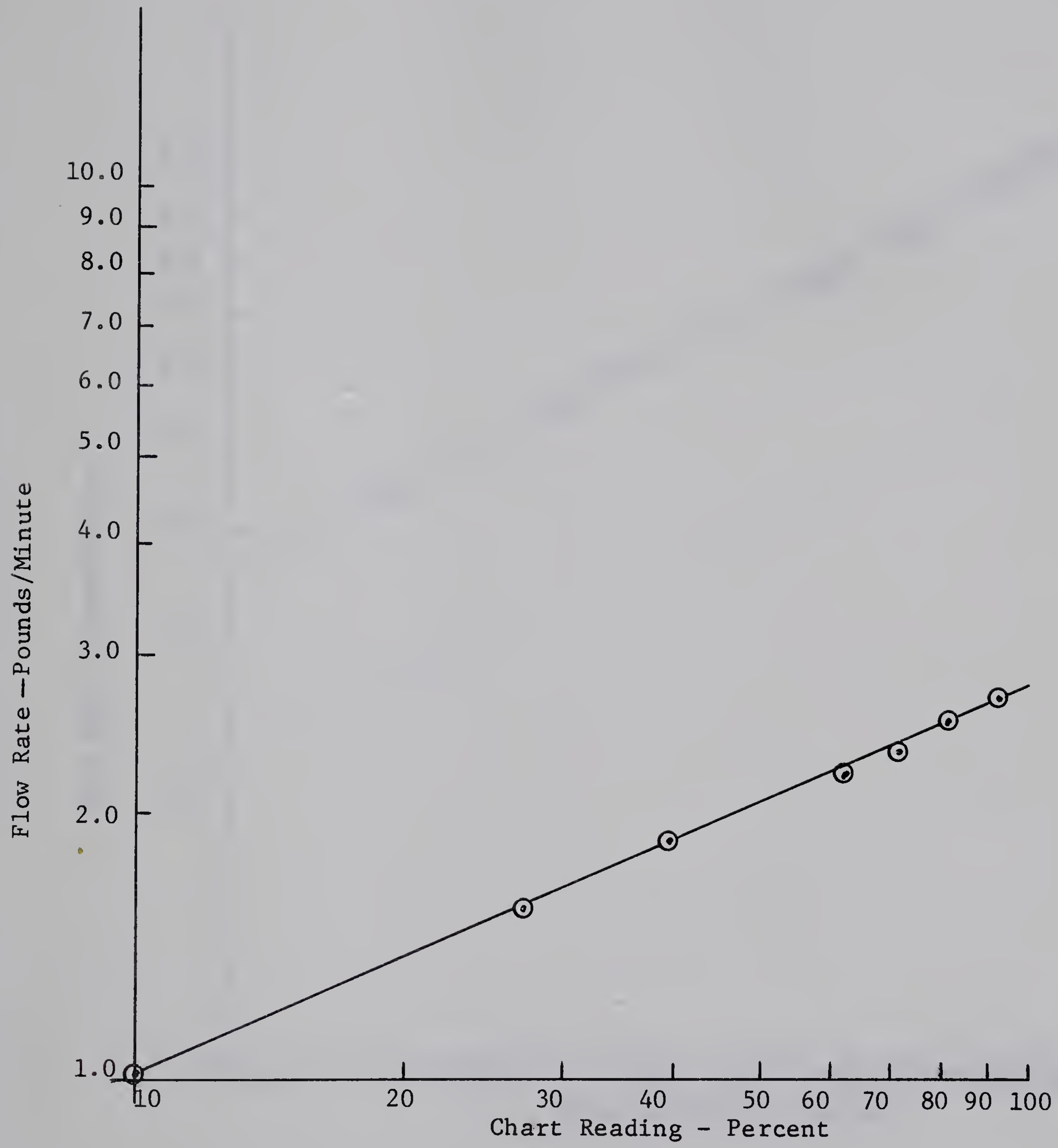


Figure B-7 CALIBRATION: SECOND EFFECT PRODUCT FLOW

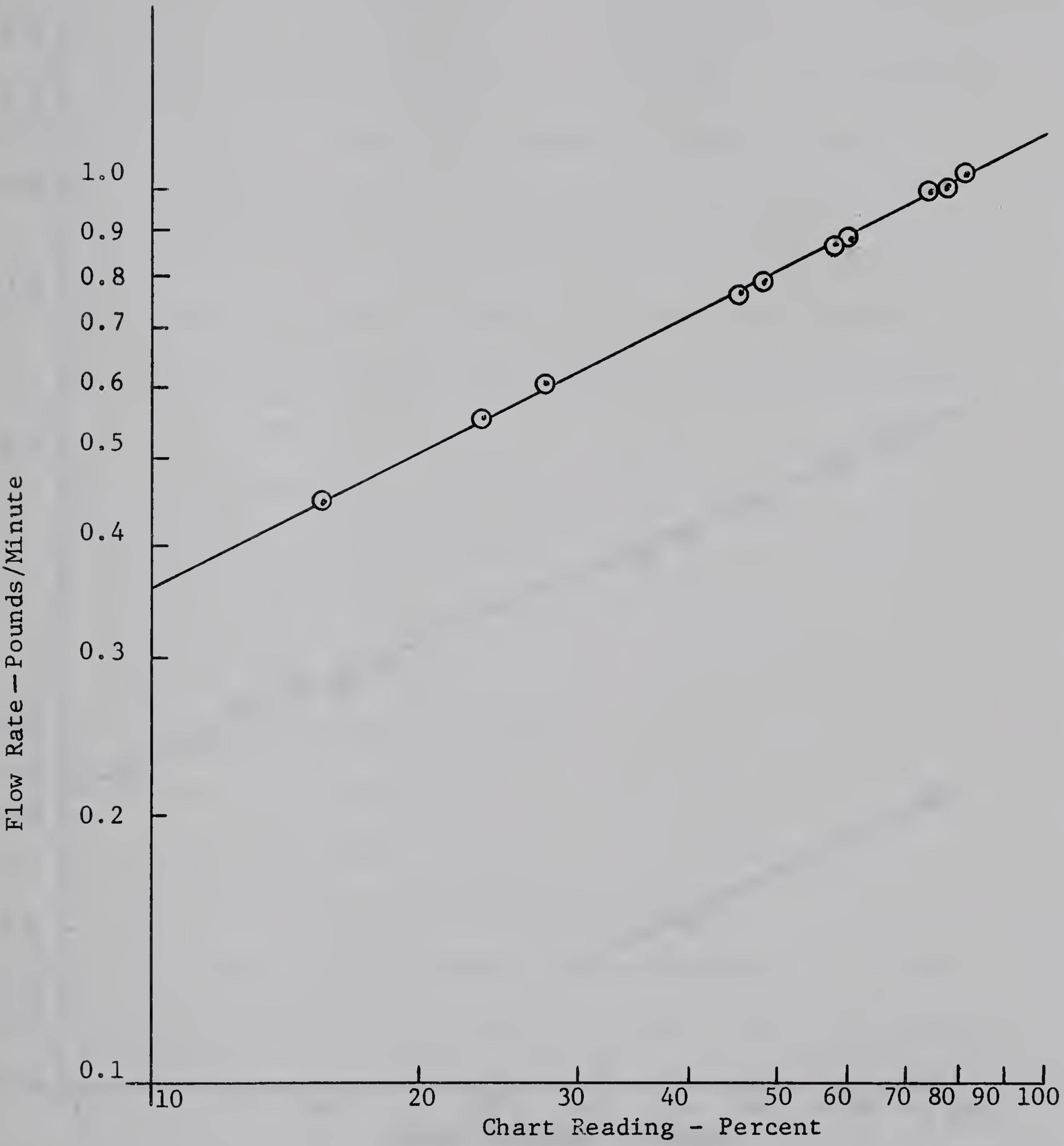


Figure B-8 CALIBRATION: FIRST EFFECT OVERHEAD FLOW (CONDENSATE)

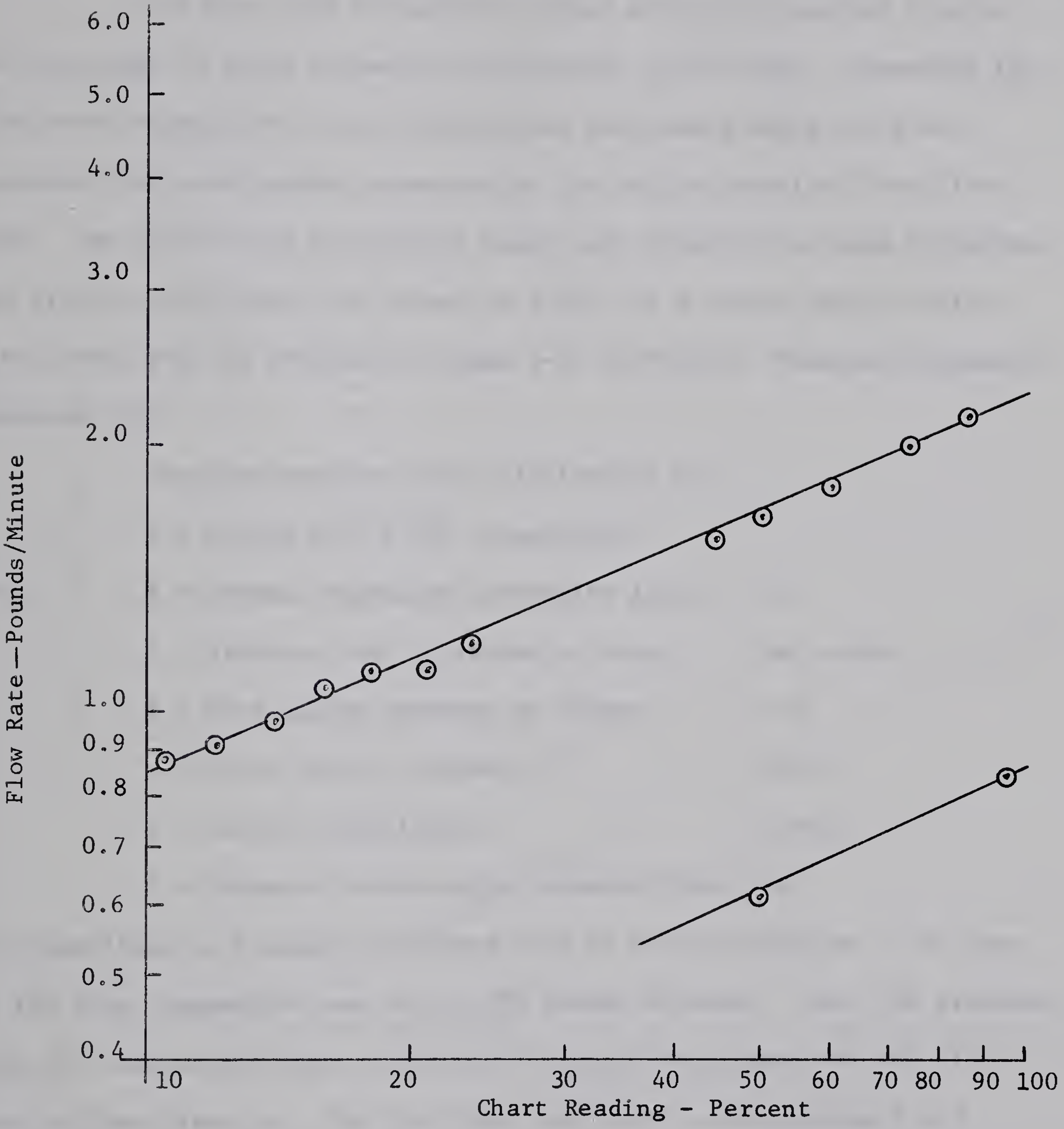


Figure B-9 CALIBRATION: CONDENSER CONDENSATE FLOW

Calibration of Orifice Run for Second Effect Circulation Line

The flow rate through the second effect circulation line is far too large to allow volumetric calibration of the flow. Therefore the flow rates through the second effect were calculated using the flow-pressure drop relationship presented in the Taylor Technical Data file (23). The orifice run and orifice plate were sized by the same procedure. The orifice coefficient was chosen as 0.625 for a square edged orifice with corner taps as presented on page 5-11 in Perry's "Chemical Engineers' Handbook" (15).

The flow-pressure drop relationship is:

$$W = 0.0997\ EK\beta^2\ D^2\sqrt{h\rho}\ \text{pounds/sec.}$$

Where E = thermal expansion correction factor (1)

h = pressure drop in inches of water (see below)

D = pipe inside diameter in inches (1.5)

ρ = fluid density pounds/ft.^{3} (60.5)

K = orifice coefficient (0.625)

β = diameter ratio: orifice diameter/pipe I.D.

The quantities in brackets are those used in the calculations. The span on the flow transmitter was set at 200 inches of water. Thus the pressure drop for various pen chart positions is easily determined and the flow rate follows directly. The flow rates and their corresponding fluid velocities (boiling ignored) were calculated as follows:

Chart Position-%	Flow lb./min.	Heating Tube Velocity ft./sec.
55	234	5.20
72	268	5.95
80	282	6.25
20	141	3.13

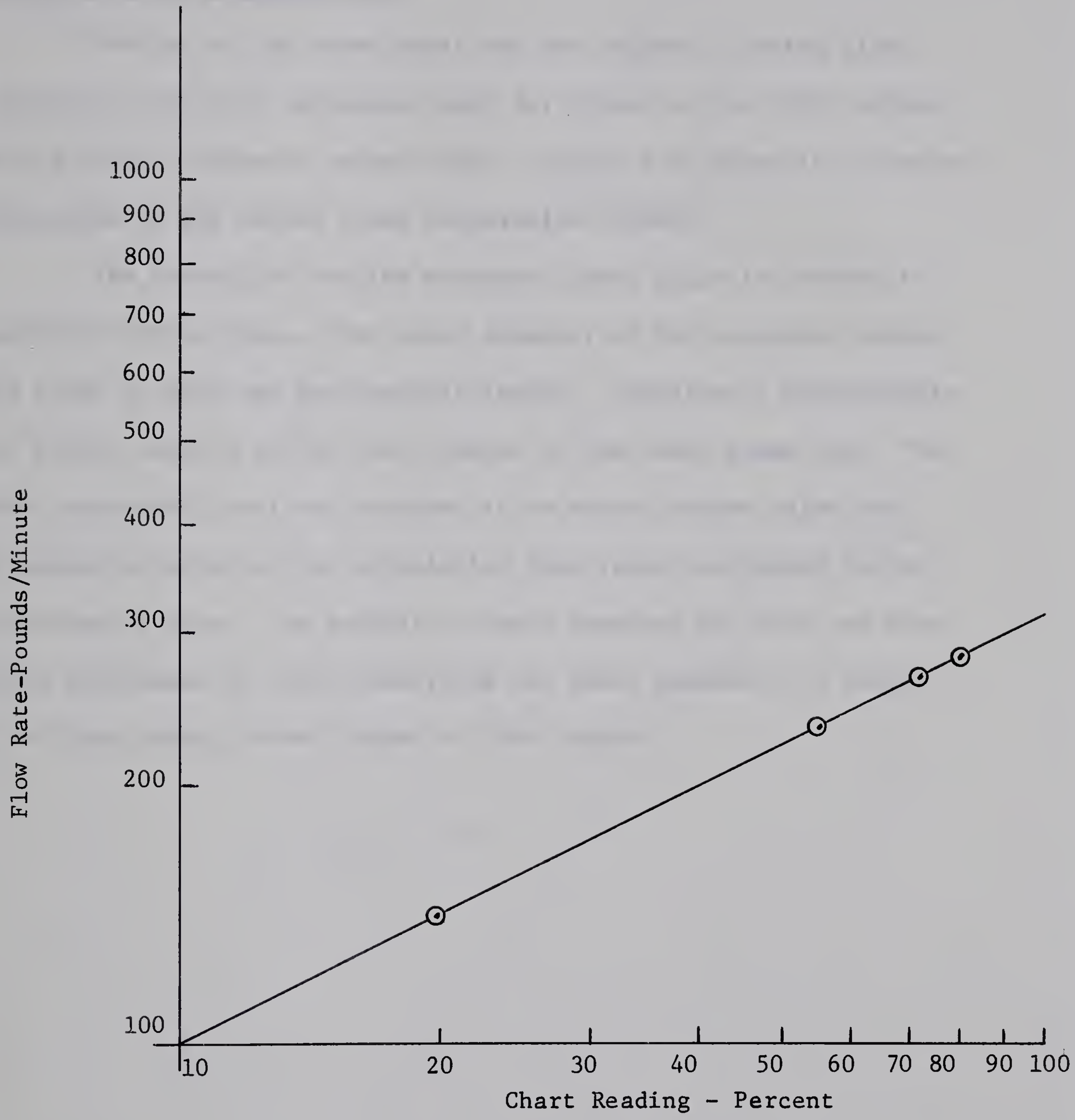
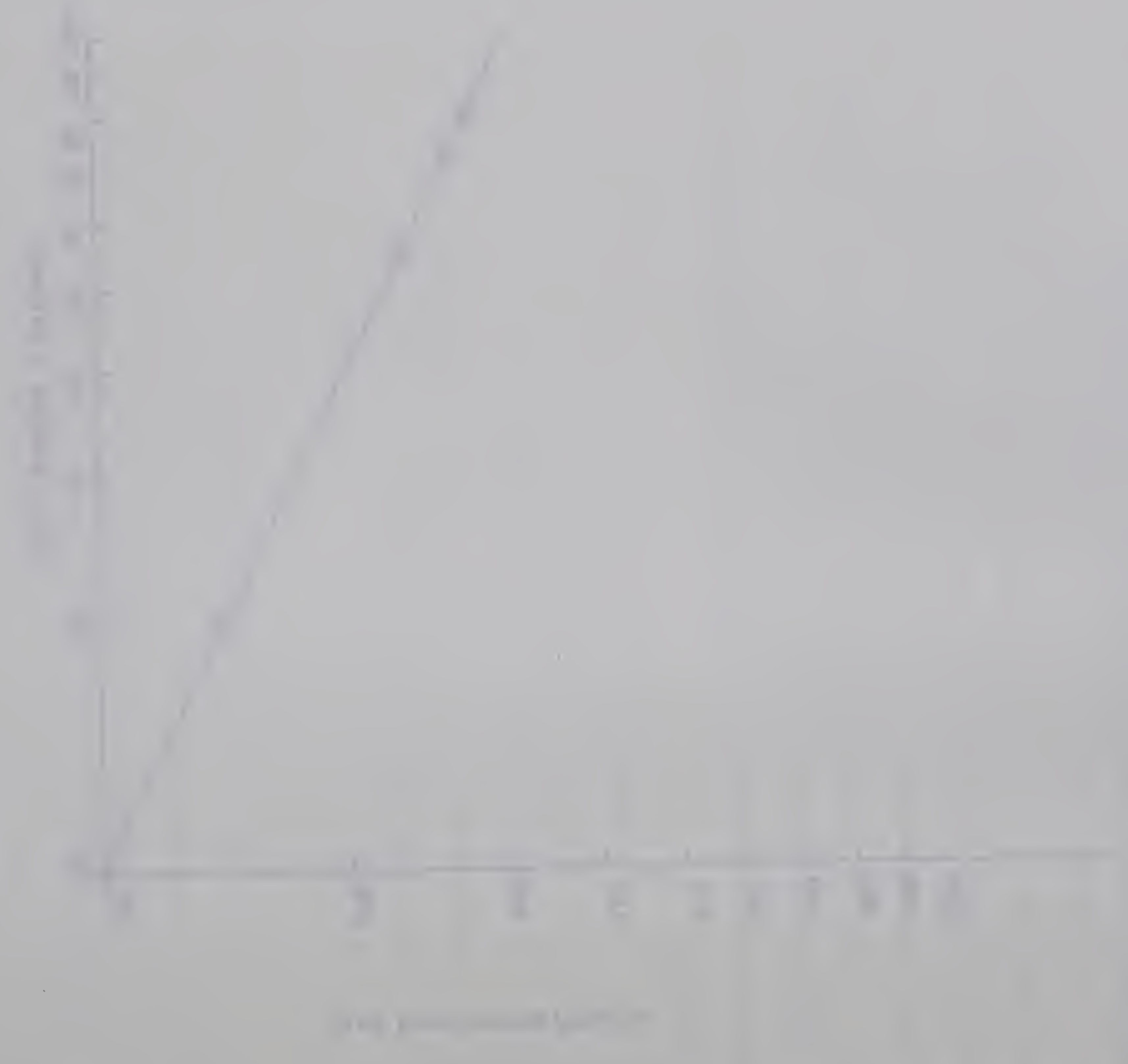


Figure B-10 CALIBRATION: RECIRCULATION FLOW



Calibration of Level Transmitters

The top of the steam chest and the separator bottom plate were choosen as the zero reference level for liquid in the first effect and second effect separator respectively. Figure B-11 shows the location of these zeros on the actual level calibration curves.

The bottom tap for the separator level guage is located in the separator bottom plate. The outer diameter of the separator outlet at this point is only one and one-half inches. Therefore a considerable venturi effect results as the fluid rushes by the level guage tap. The indicated separator level was measured to be eleven inches below the actual separator level at the circulation flow rates maintained during the experimental runs. The separator levels reported for each run were therefore determined by first converting the chart reading to a liquid level and then adding eleven inches to that number.

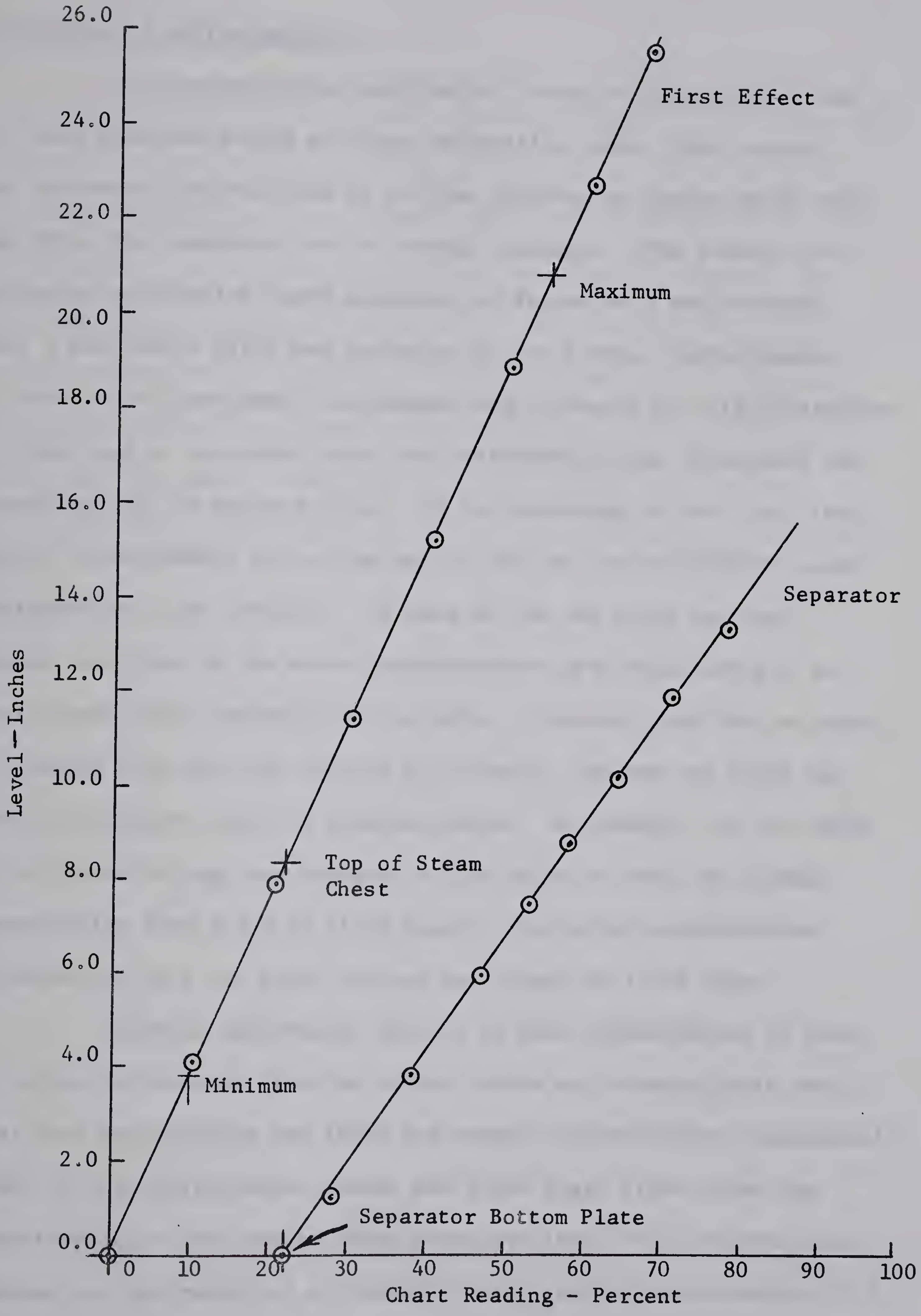


Figure B-11 CALIBRATION OF LEVEL TRANSMITTERS

Calibration of Refractometers

The concentration calibration curves of figures B-12 and B-14 were obtained during off-line calibration runs. Both curves were re-checked and verified by on-line calibration checks which were made while the evaporator was in normal operation. The product concentration calibration curve presented in Figure B-13 was obtained after a new source light was installed in the product refractometer. The span of the instrument was changed significantly by this alteration but could not be corrected since the refracometer span adjustment was already set at its minimum value. At the beginning of each run, the product refractometer output was set to fall on the calibration curve programmed into the computer. As long as the set point was not changed, an error in the actual refractometer span would not put any final steady state concentration in error. However, when the set point was changed with the span setting not correct, the new set point was actually different from its intended value. For example, in run 16FFB the set point voltage was changed to 4.50 volts to take the product concentration from 8.35% to 11.0% sugar. The actual concentration equivalent of this set point voltage was closer to 12.0% sugar.

Another unfortunate feature of this refractometer is that its output drifts away from the correct value at an appreciable rate. After four hours during run 16FFB the product concentration represented by 86% of the refractometer output was 0.30% sugar higher than the concentration at that output three hours earlier. This concentration increase was the result of a downward drift rate of approximately 0.1% sugar/hour in the refractometer output. Since this drift is recognized

as an error by the feedback control loop, a correction to the steam flow was made by the controller to eliminate this error. When the product concentration readings for all experimental runs using feedback control were plotted, it was therefore necessary to adjust the product concentration by subtracting the drift error from the measured concentration. This adjustment brought the product concentrations back to their set point value at the final steady state condition. Quite clearly, any stable feedback control system with integral control action must return the controlled variable to its set point following a disturbance.

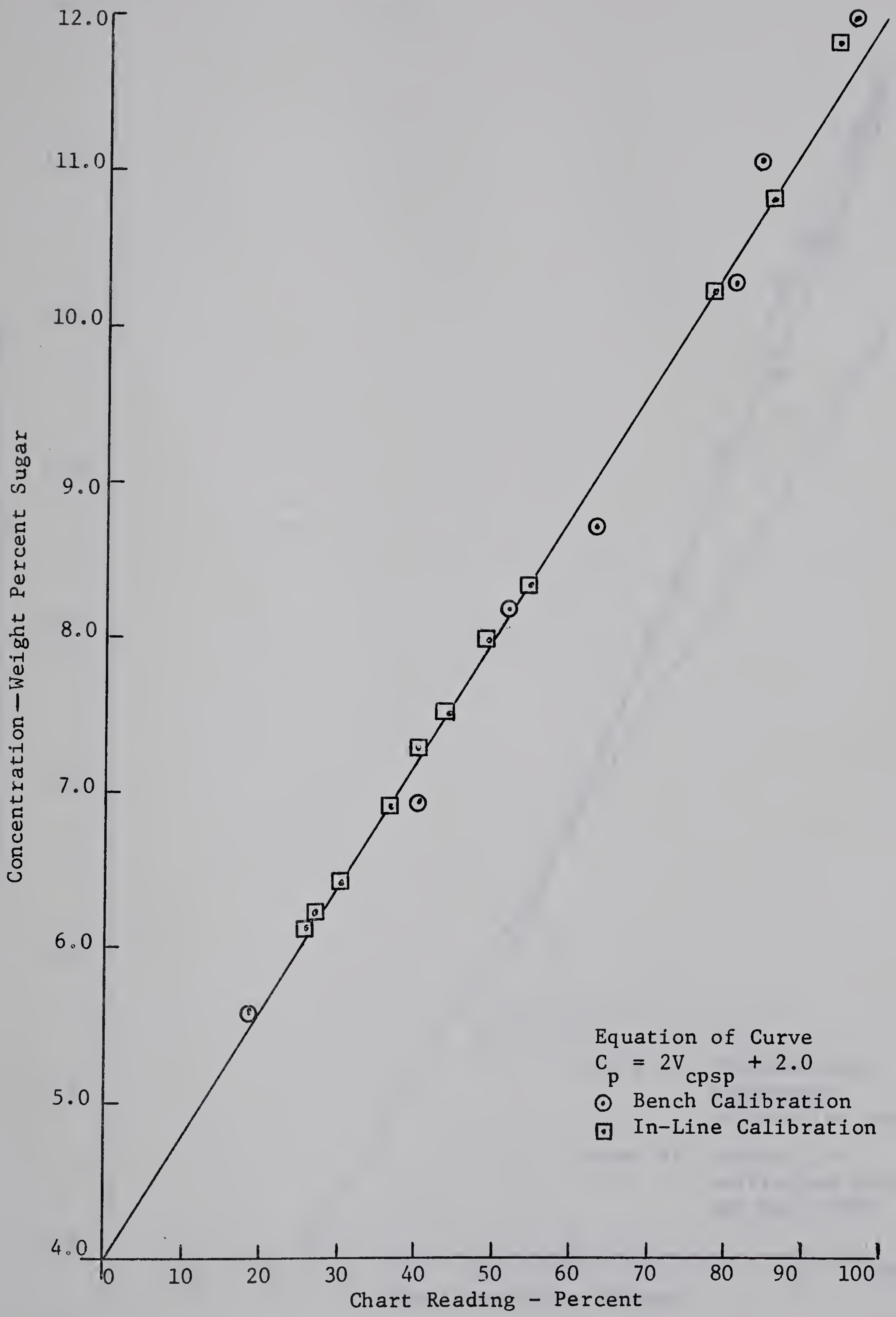


Figure B-12 CALIBRATION: PRODUCT COMPOSITION

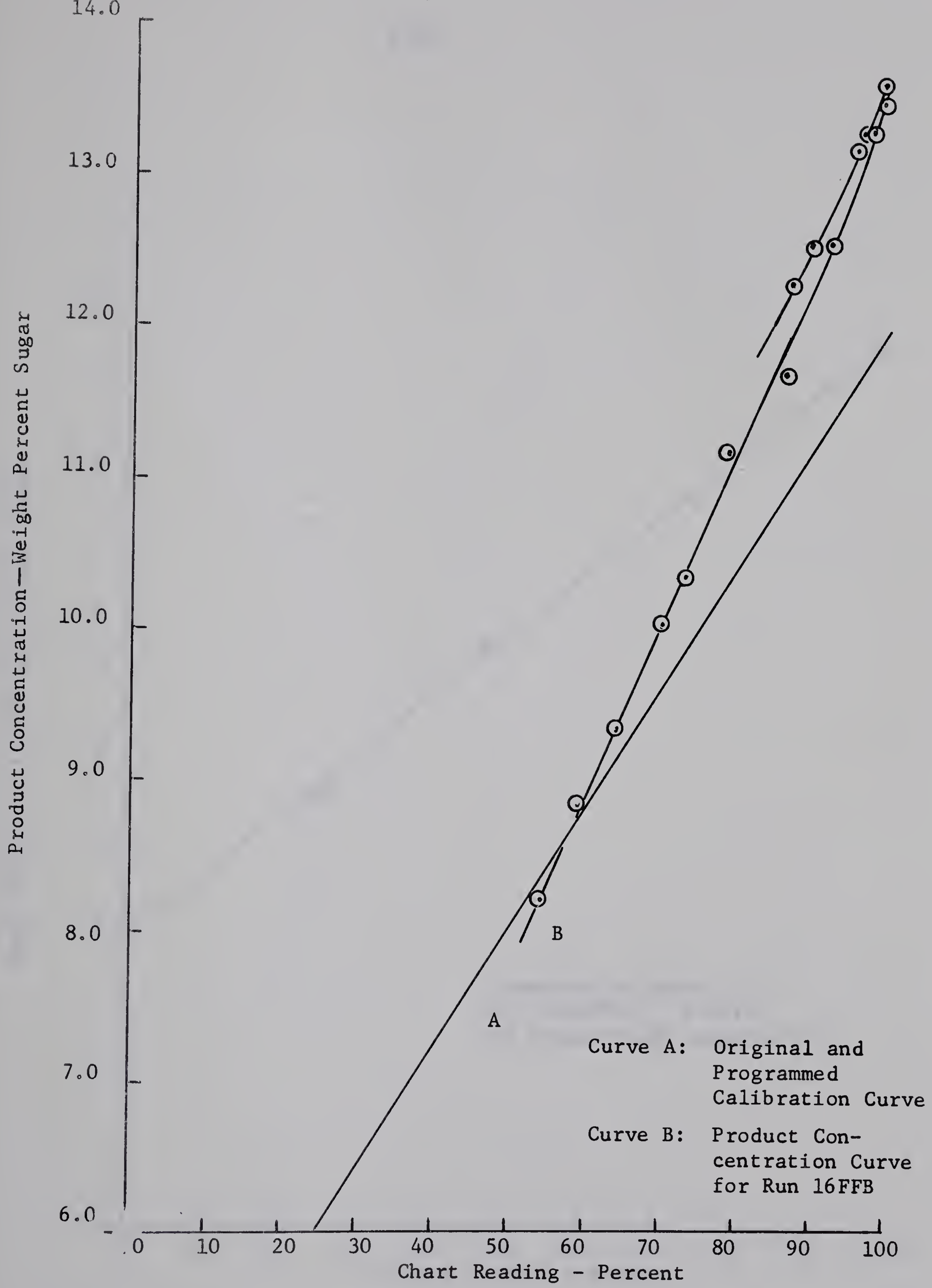
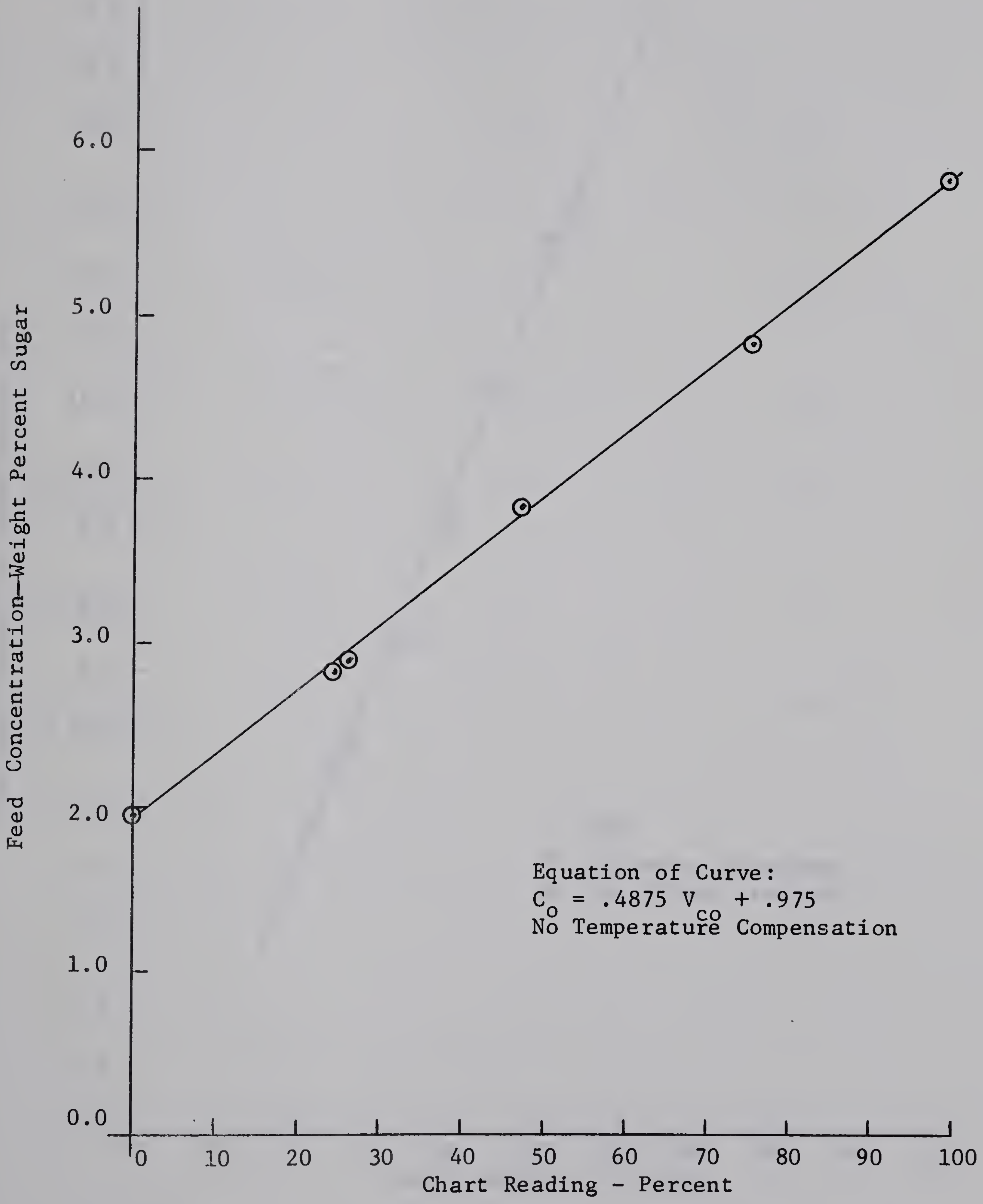


Figure B-13 CALIBRATION: PRODUCT CONCENTRATION FOR RUN 16FFB



Equation of Curve:
$C_o = .4875\ V_{co} + .975$
No Temperature Compensation

Figure B-14 CALIBRATION: FEED CONCENTRATION

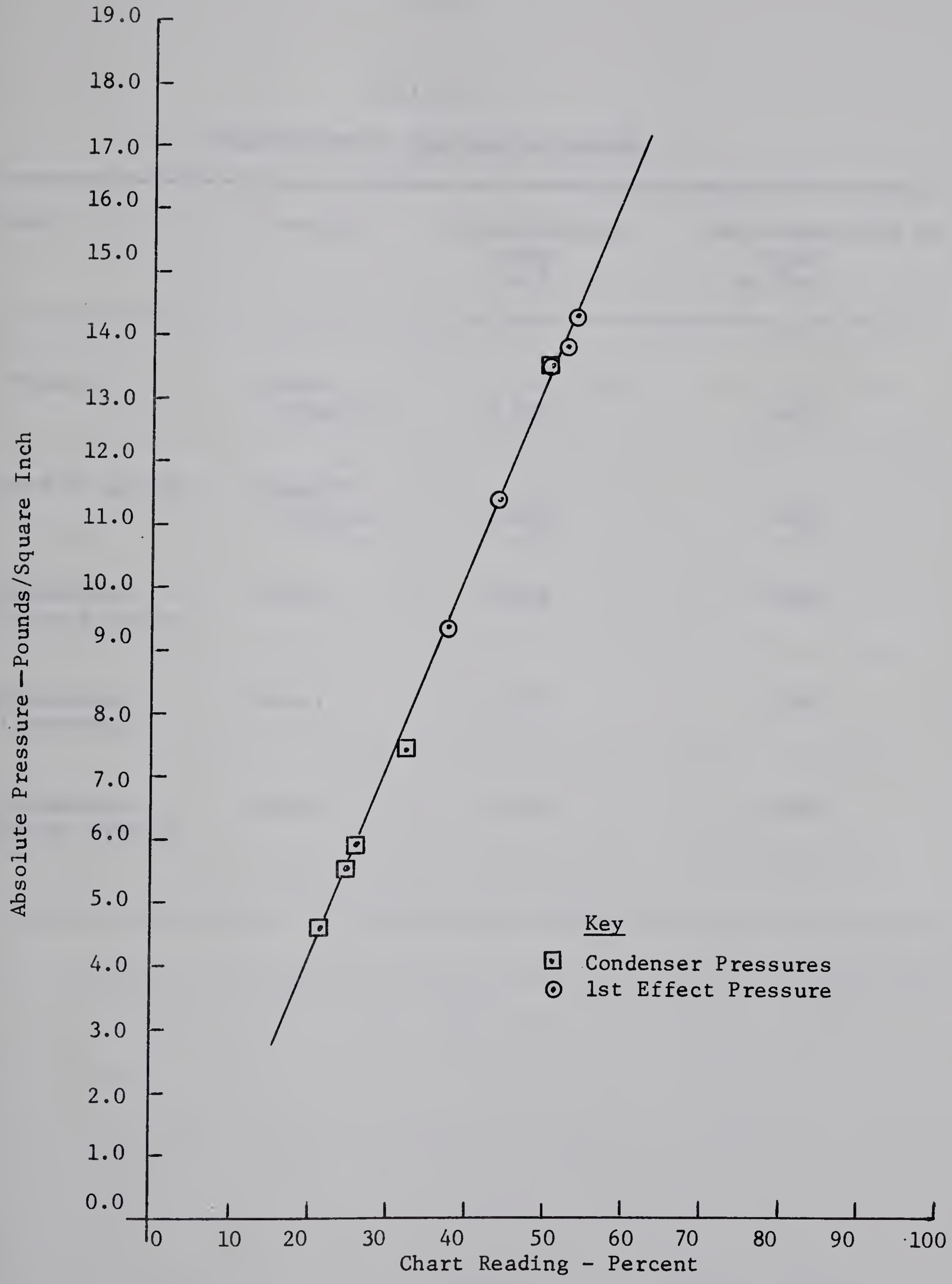


Figure B-15 CALIBRATION OF PRESSURE TRANSMITTERS

Table B-2

CALIBRATION OF EVAPORATOR TANKAGE

Tank	Contents	Cross-Sectional Area $ft.^2$	Tank Volume/Inch of Level $ft.^3/in.$
Product	Sugar Solution	3.140	.2616
Feed #1 and #2	Sugar Solution	4.898	.4081
Condensate (Second Effect)	Water	4.898	.4081
Condensate (Condenser)	Water	1.319	.1099
Condensate (First Effect)	Water	1.319	.1099

APPENDIX C

FEEDFORWARD PRODUCT CONCENTRATION CONTROLLER

A. Analog Circuit

The basic analog circuit uses the measured process variables W_o and C_o and the internally set variables C_{psp} and K_s to compute the steady state steam flow which will provide product of set point concentration. Additional steam requirements are computed and added to the material balance steam flow to supply latent heat to the feed and to offset heat losses from the evaporator. Negative concentration feedback control action can be added to the computer circuit by adding the corrective action of the Foxboro concentration controller to the analog circuit output voltage.

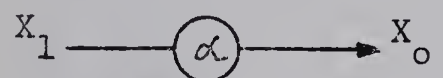
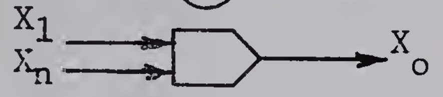
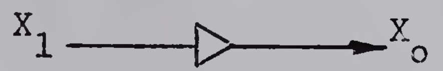
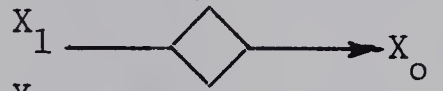
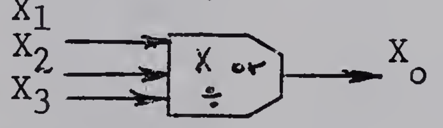
Figure C-1 outlines the circuit which computes the steam flow controller set point voltage V_s. Most of the symbols and nomenclature used in Figure C-1 have been previously defined; however, the few following in Table C-1 do merit summary here.

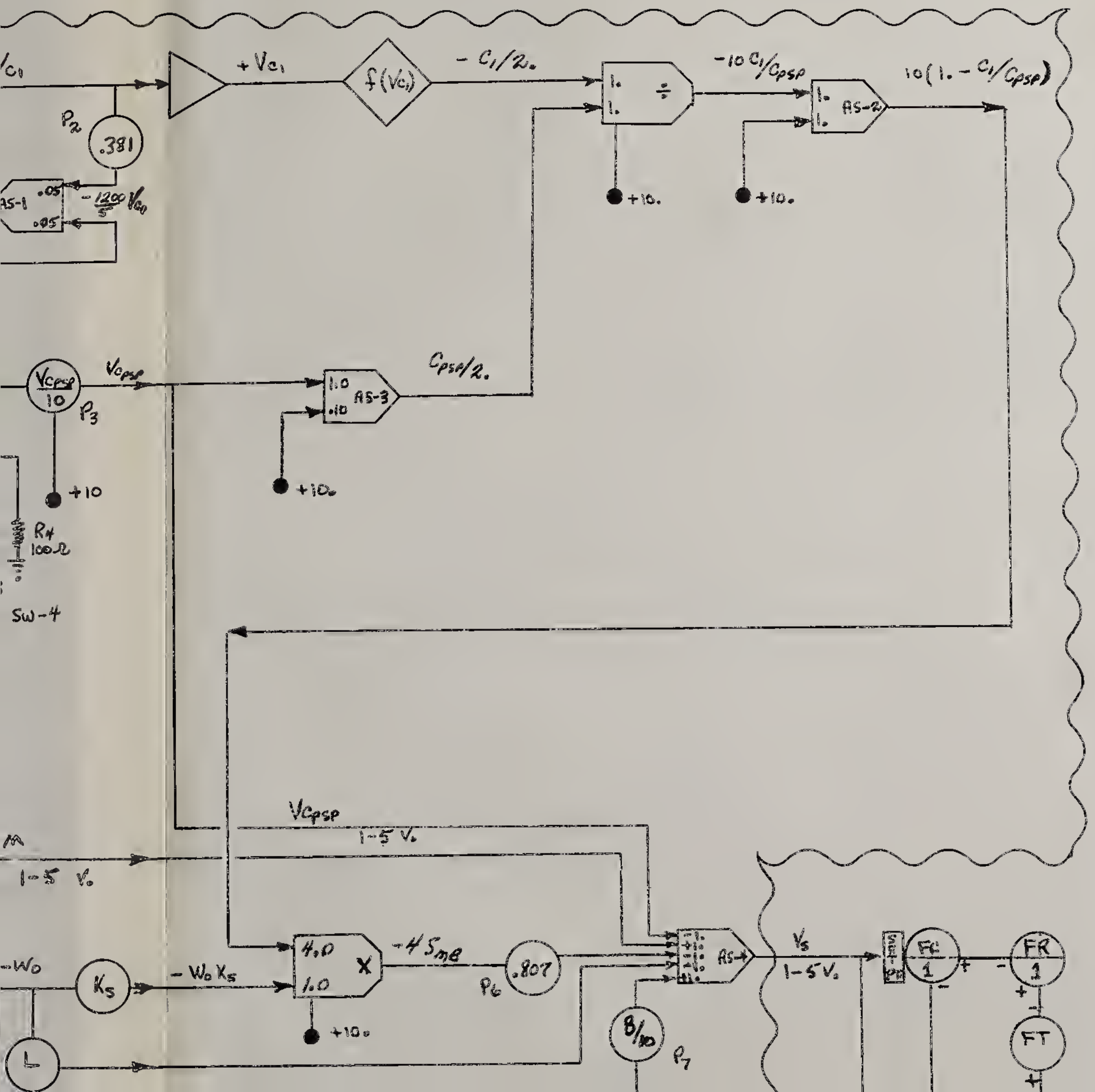
B. Interfacing the Analog Control Circuit with Foxboro Instrumentation

Some difficulty was encountered in transfering voltage signals between the Foxboro instrumentation and the analog controller before the system depicted in Figures C-1 and C-2 was developed. These signals were either in the 1 to 5 volt or 2 to 10 volt range and were obtained as the voltage differential across 100 ohm (R_4, R_5, R_6) or 200 ohm resistors (R_1, R_2, R_3) respectively. In order that the differential signals could be transmitted without error, as absolute

Table C-1

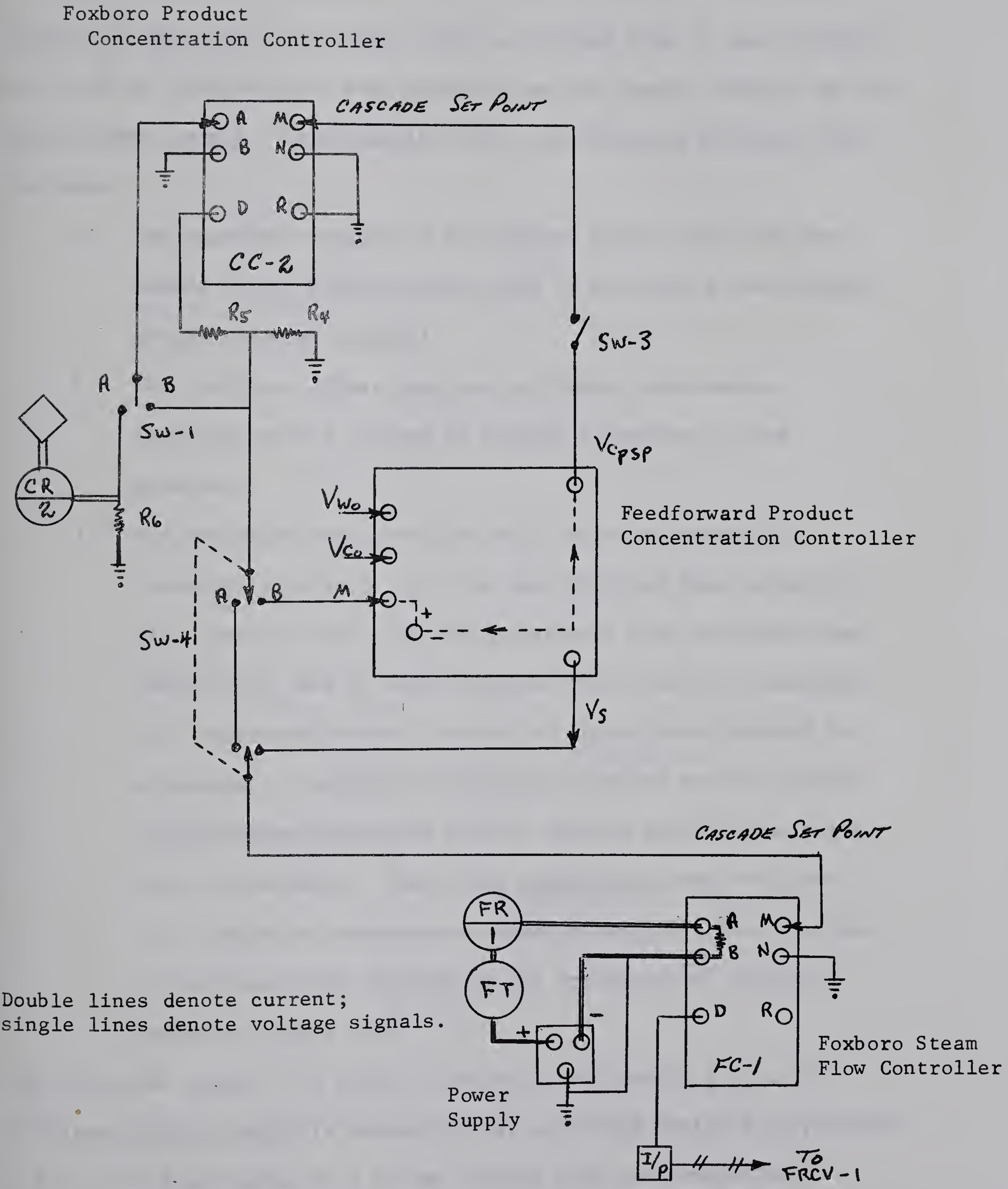
SPECIAL SYMBOLS USED IN FIGURE C-1

Symbol	Name	Value
M	Foxboro Concentration Controller Output (Volts)	$M = K_c(V_{cpsp} - V_{cp}) + K_c\int(V_{cpsp} - V_{cp})dt + V_{cpsp}$
L	Latent Heat Term Volts/lb. feed (scaled)	$L = 3.226\ (T_1 - T_o)/\Delta H_s$
B	Steam Set Point Bias (Volts) (Scaled)	$B = -1.774 + 3.226$ (Bias Steam Flow)
S	Total Steam Flow (lb./min.)	$S = \underbrace{W_oK_s(1-C_1/C_{psp})}_{\text{Steam Flow}} + \frac{W_o\Delta t}{\Delta H_s} + \text{Bias}$
V_s	Steam Flow Set Point (Volts)	$V_s = 3.226S - 1.774 + K_c(V_{cpsp} - V_{cp}) + K_c\int(V_{cpsp} - V_{cp})dt$
	Adjustable Coefficient $0 \le \alpha \le 1$	$X_o = \alpha X_1$
	Add - Subtract Unit	$X_o = X_1 \pm X_n \quad n = 2$ to 6
	Amplifier (Sign Changer)	$X_o = -X_1$
	Adjustable Function Segment	$X_o = mX_1 + b$
	X Multiplier	$X_o = X_1\ X_2/10_1$
	÷ Divider	$X_o = 10\ X_1/X_3$



Vc1
+Vc1
f(Vc1)
- C1/2.
1.
1.
÷
+10.
-10 C1/Cpsp
1.
1.
AS-2
+10.
10(1. - C1/Cpsp)
P2
.381
AS-1
.05
.05
- 1200/5 Vc0
Vcpsp/10
P3
Vcpsp
+10
R4
100Ω
SW-4
1.0
.10
AS-3
+10.
Cpsp/2.
Vcpsp
1-5 V.
1-5 V.
-W0
Ks
- W0 Ks
4.0
1.0
X
+10.
-4 SmB
.807
P6
8/10
P7
AS-4
Vs
1-5 V.
FC
1
FR
1
FT
L

[illegible]



Double lines denote current;
single lines denote voltage signals.

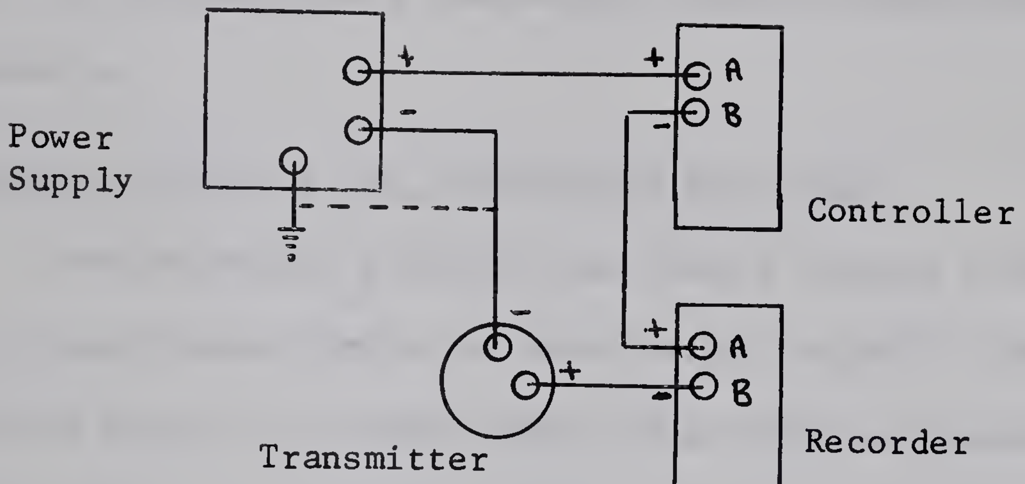
Figure C-2 INTERFACING OF FOXBORO CONCENTRATION AND STEAM FLOW CONTROLLERS WITH ANALOG FEEDFORWARD CONCENTRATION CONTROLLER

values, to the analog controller, the low voltage side of each resistor had to be at precisely the same potential as the common terminal on the module power supply. To accomplish this, the following procedure was followed:

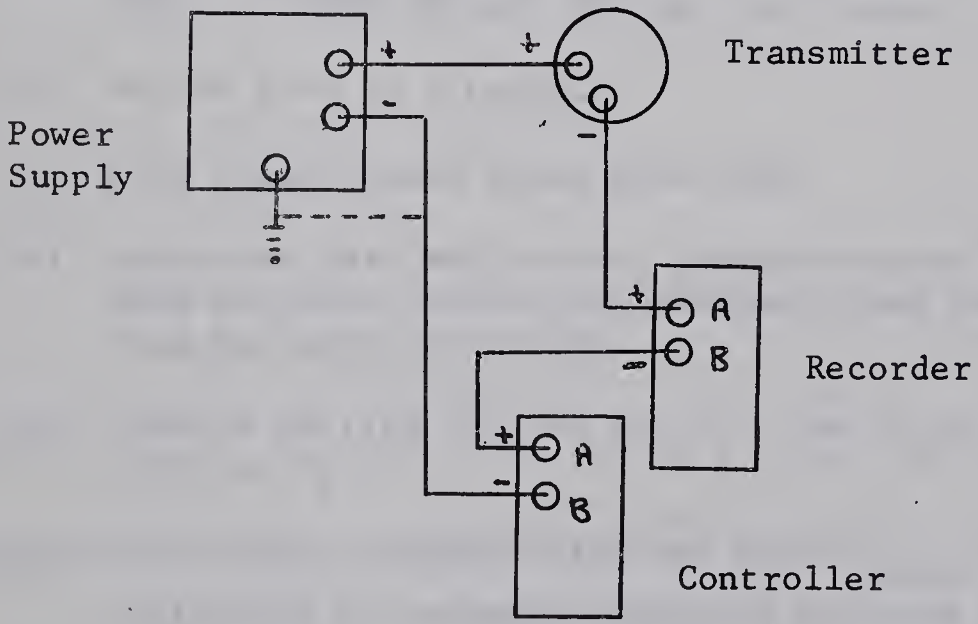
1. The negative terminal of the Foxboro power supply was connected to its ground terminal and to the module power supply ground (common) terminal.
2. The individual power supplies in Foxboro instruments (controllers and voltage to current transducers) were grounded.
3. The resistors were installed with their low potential terminals at ground. For the feed flow and feed concentration control loops, the desired signals were available when resistors R_1 and R_3 were installed below the last component. The "preferred" control circuit of Figure C-3A required no alteration. Both the steam flow controller and the product concentration controller were to receive set point voltages from the computer. Thus these controllers must both have their negative terminals at ground potential. This was done by switching both circuits to the "permissible" configuration of Figure C-3B.

The set point signal to a Foxboro controller is usually a 10 to 50 milliamp current, which is dropped across a 100 ohm resistor to provide a 1 to 5 volt potential to a bridge circuit with the controller.

A. Preferred Circuit



B. Permissible Circuit



Note: For both configurations the negative terminal of the power supply may be grounded.

Figure C-3 WIRING CONFIGURATIONS FOR FOXBORO CONTROL LOOPS

The 10 to 50 milliamp controller input is dropped across a similar resister (resister R_7 in Figure C-2) to enter the bridge circuit as the measured variable. The set point resistor was removed from the steam flow and product concentration controllers so that a 1 to 5 volt signal, from the feedforward controller, could be used directly by each controller.

C. Balancing Procedure for Feedforward Controller

The balancing procedure was always started with the steam flow at a steady state flow rate under manual control. The product concentration was at its steady state value which corresponded to the steam flow rate.

1. Feedforward Control

(i) Turn on power to analog controller

(ii) Switch on Sw-3 and Sw-5

(iii) Switch Sw-1 to B position and set control action on CRC-2 to 100% PB and .20 min. per repeat.

(iv) Switch Sw-4 to B position

(v) Note steady state steam flow rate

(vi) Determine feed and product concentrations by analysis, feed and first effect temperatures, and feed flow rate from pen chart positions.

(vii) Compute setting "L" for pot P_4 - Set P_4 and Set K_s = 0.500 on P_5.

(viii) Set product concentration set point V_{cpsp} to voltage calibrated to represent analyzed value of C_p.

(ix) Set V_{co} at calibration voltage which represents analyzed value for V_{co}. Leave the initial condition on the time delay circuit integrating amplifier until after the steam flow set point bias has been set (Step xi)

(x) When Foxboro concentration controller output equals the set point voltage V_{cpsp}, switch the Foxboro steam flow controller to its cascade set point position.

(xi) Note steam flow: if it changes at all, rapidly reset bias pot P_7 to bring steam flow back to the value noted in v above.

(xii) Remove initial condition on integrating amplifier.

2. Feedforward - Feedback Control

The procedure used in this case is identical to that outlined above, up to and including step (xi). At this point the steam flow is the correct steady state value for operating conditions as they exist during the balancing procedure. Following step xi, simply switch Sw-1 to position A and remove the initial condition on the integrating amplifier. This places the actual product concentration as the measured value for the Foxboro concentration controller and in effect adds feedback control to the control system which presets the steam flow rate.

3. Feedback Control

Since the product concentration controller receives a cascade set point voltage, the manual-to-automatic balancing mechanism within the controller was not available to help balance the steam flow when the unit was being switched to feedback control.

The procedure therefore was to determine the product concentration by analysis and set the product concentration set point V_{cpsp} at that value. The pen position was reset if necessary. When Sw-4 was switched to position A and the steam behavior was noted.

If deflection from the steady state flow resulted, the reset rate was quickly increased to overcome reset windup and bring the controller to its correct value again. In most cases a trial and error procedure was required before the steam flow could be held steadily on its correct value with the controller settings which were to be used during the experimental run.

When many feedback runs are to be made it is recommended that the computer supplied set point be disconnected from the Foxboro controller, the cascade jumper should be removed from the controller, and the set point should be adjusted on the controller. Balancing the controller output before switching from manual control to automatic control will then be a very simple matter.

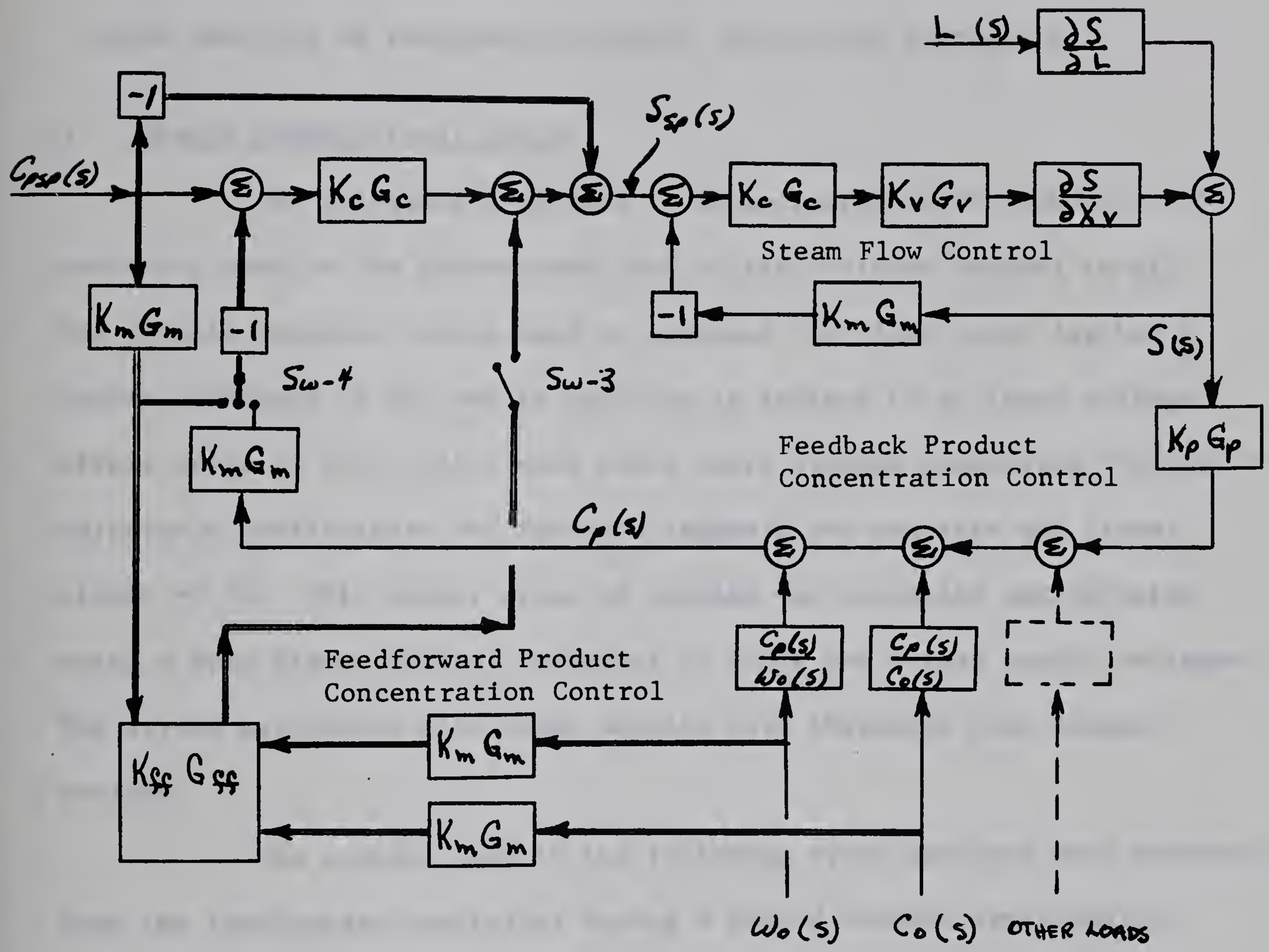


Figure C-4 BLOCK DIAGRAM OF PRODUCT CONCENTRATION CONTROL SYSTEM.

Nomenclature of Figure C-4.

K	Gain
G	Transfer Function
L	Load
X	Valve Stroke

C_p, C_o, W_o, & S : Process Variables as Previously Defined.

Subscripts:

c	Controller
v	Valve
p	Process
m	Measurement
sp	Set Point
ff	Feedforward Controller.

APPENDIX D

ERROR ANALYSIS OF FEEDFORWARD PRODUCT COMPOSITION CONTROLLER

A. Module Computational Errors

The tolerance specified by Consolidated Electrodynamics for resistors used on the add-subtract and multiply-divide modules is ±1%. The dynamic response module used to generate the first order lag has a poorer tolerance of ±5% and in addition is subject to an input voltage offset of up to 0.01 volts, with 0.003 volts offsets considered "typical." Adjustable coefficients and function segments are accurate and linear within ±0.3%. This latter group of modules was installed and adjusted using a four figure digital voltmeter to check the actual output voltages. The errors associated with these modules have therefore been largely avoided.

The numbers used in the following error analysis were obtained from the feedforward controller during a period between experimental runs. The controller inputs were taken directly from the operating evaporator in the manner illustrated in Figure C-1. All voltage readings were obtained using the digital voltmeter.

Errors are generally presented as absolute numbers. Where they are computed as percentages, they are based on the theoretically correct output from each module. All numbers are voltages unless indicated otherwise. The analysis of each module and the error associated with it are as follows:

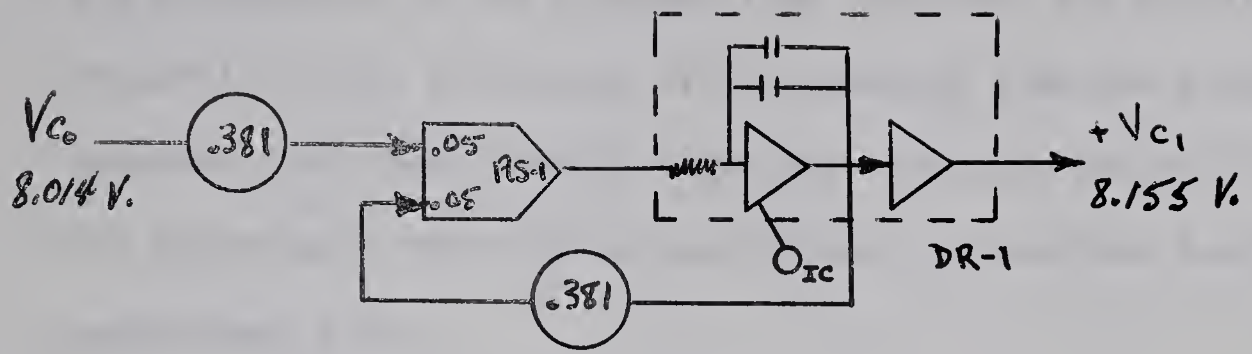
SYSTEM INPUTS:

$V_{co} = +8.014$ v.

$V_{wo} = +4.370$ v.

$V_{cpsp} = +3.100$ v.

1. Time Delay Circuit



a.	To P_1: +8.014	From P_1:	+3.054
		Theortetical Output	+3.053
		Error	+0.001
b.	To AS-1: -3.105	From AS-1:	0.000
	+3.054	Theoretical Output	-0.051
		Error	-0.051
c.	To DR-1: 0.000	From DR-1:	-8.155
	Grid Offset = 0.000	Theoretical Output	-8.014
		error	0.141
d.	To P_2: -8.155	From P_2	-3.105
		Theoretical Output	-3.107
		Error	-0.002

Circuit Error = .141 = 1.73%

e. Accuracy of first order lag.

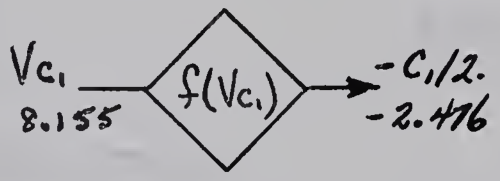
The first order time delay was tested using a five minute time constant (pots P_1 and P_2 were set at .762) A step increase and a step decrease in V_{co} and the delayed response of V_c were plotted. As accurately as could be determined from the plots, the results were:

Step Change:	Response 63.2% Complete:
V_{co} 5.000 to 9.000	5.25 Minutes
V_{co} 8.000 to 5.000	4.80 Minutes

Average Time Constant: 5.025 minutes.

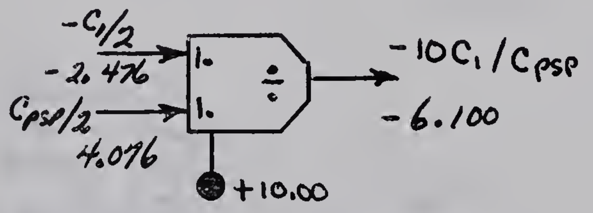
The differences in the observed time constants are attributed primarily to the difficulty of interpreting from the plotted response, the exact time at which each response was 63.2% complete. The approximate 5% error in each figure is therefore basically a measurement error.

2. Feed Composition Function Segment



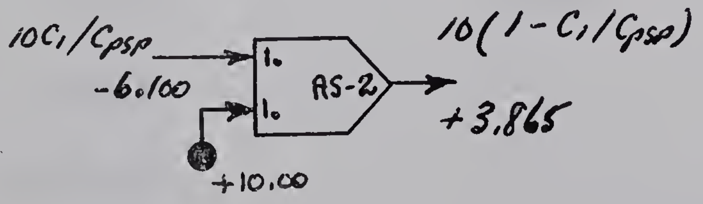
To $f(V_c)$: 8.155		From $f(V_c)$	-2.476
		Theoretical Output	-2.475
Function: $C_1/2 = .2438\ V_{c1} + .487$		Error	-0.001

3. Divider



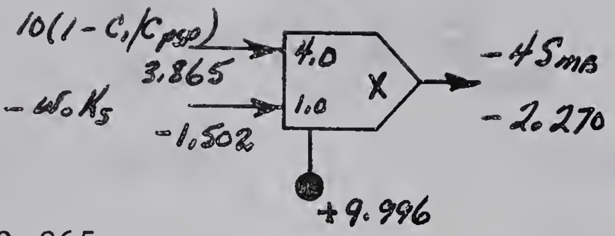
To ÷ :	-2.476	From ÷ :	-6.100
	+4.076	Theoretical Output	-6.075
	+10.00	Error	-0.025

4. Add-Subtract Unit AS-2



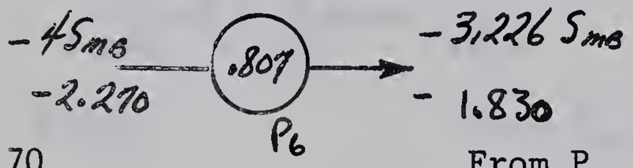
To AS-2:	-6.100	From AS-2 :	+3.865
	+ 10.00	Theoretical Output	+3.900
		Error	-0.035

5. Multiplier



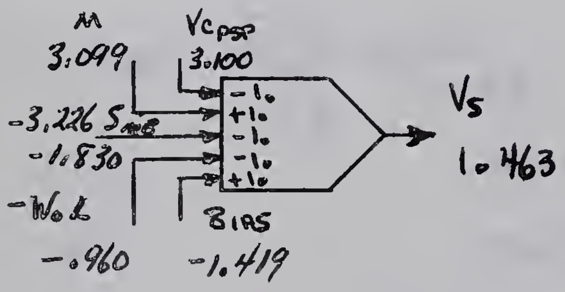
To X:	+3.865	From X :	-2.270
	-1.502	Theoretical Output	-2.322
	+9.996	Error	+0.052

6. Adjustable Coefficient P_6



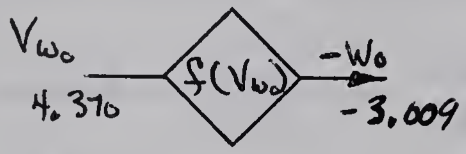
To P_6 :	-2.270	From P_6	-1.830
		Theoretical Output	-1.832
		Error	-0.002

7. Add-Subtract Unit AS-4



To AS-4 :	-(-1.830)	From AS-4	+1.463
	-(-0.960)	Theoretical Output	1.370
	-1.419	Error	-0.093
	+3.099		
	-(+3.100)		

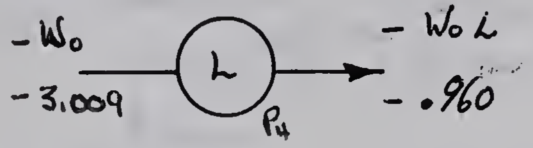
8. Feed Flow Function Segment $f(V_{wo})$



To $f(V_{wo})$:	+4.370	From $f(V_{wo})$	-3.009
		Theoretical Output	-2.999

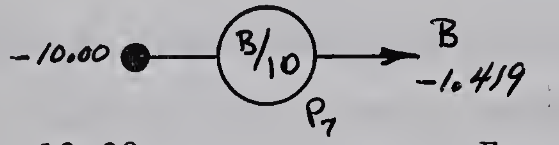
Function: $W_o = 0.5375\ V_{wo} + 0.655$ Error   -0.010

9. Adjustable Coefficient P_4 (Latent Heat)



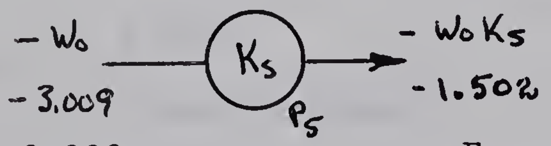
To P_4 :	-3.009	From P_4 :	-0.960
L = .319		Theoretical Output	-0.960
		Error	0.0

10. Adjustable Coefficient P_7 (Bias)



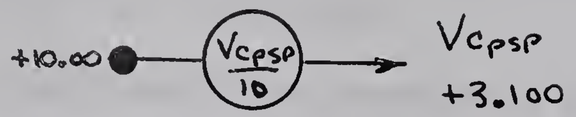
To P_7 :	-10.00	From P_7	-1.419
B =	.1419	Theoretical Output	-1.419
		Error	0.0

11. Adjustable Coefficient P_5 (K_s)



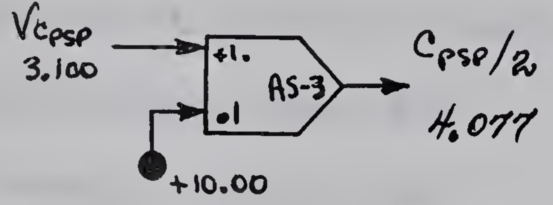
To P_5 :	-3.009	From P_5	-1.502
K_s =	.500	Theoretical Output	-1.504
		Error	-0.002

12. Adjustable Coefficient P_3 ($V_{cpsp}/10$)



To P_3 :	+ 10.00	From P_3 :	+3.100
		Theoretical Output	+3.100
		Error	0.0

13. Add-Subtract Unit AS-3



To AS-3 :	+3.100	From AS-3	+4.077
	+10.00	Theoretical Output	+4.100
		Error	-0.023

B. Theoretical Controller Output Voltage V_s

The theoretically correct value for the steam set point voltage can be obtained by performing the same calculations on the input variables as are done by the computer. The system voltage inputs previously listed represent:

C_o = 4.882%

C_{psp} = 8.200%

W_o = 2.999 lbs./min.

The actual steam flow must be adequate to supply latent heat to the feed, replace heat losses and to concentrate the feed solution to the set point concentration. For the error analysis the parameters which determined the steam requirements of each division above were:

Steam Economy K_s = .500

Latent heat steam (with its conversion factor included)

= .319 v. / lb. feed

Heat loss bias (with conversion bias included) = 1.348 v.

Although the input bias to add-subtract unit AS-4 was -1.419 volts, a low precision input resistor reduced the gain on this term to below 1.0. This input bias, when isolated from the other inputs to AS-4, gave an output signal of -1.348 volts as indicated above.

The calculated steam requirements, after conversion to the steam flow set point voltage signal are:

(a) material balance steam : 1.959 volts
(b) latent heat steam : .957 volts
(c) bias : -1.348 volts

V_s = 1.568 volts.

The computer output of 1.463 volts is therefore .105 volts or 6.70% below its theoretical value.

In practice the errors inherent in the feedforward composition controller components were avoided by the steam flow balancing precedure outlined in Appendix C. Thus in this instance, if a set point signal of 1.568 volts was necessary to maintain the product composition at 8.20% sugar, the bias voltage would have been increased by .105 volts so that the output of AS-4 reached 1.568 volts.

C. Operational Errors

1. Integrator grid offset voltage error When the previous error analysis was made, the initial condition was held on the integrator of module DR-1 to prevent the input concentration signal from changing

while measurements were being made. During experimental runs this was not done and an additional source of error resulted due to the offset in grid voltages encountered at the integrator. Table D-1 summarizes the time delay circuit offsets and the resulting errors in the feed composition for six of the experimental runs. If is evident that the grid offset voltage was usually very close to the typical figure of 0.003 V.

Due to the very small loop gain below the integrator output the concentration voltage is reduced by a factor of 0.019. Thus the concentration error that would produce an offset of 0.003V is 0.003/.019 = 0.157 volts. This is the same order of magnitude as the observed concentration offsets. For the last three runs in Table D-1, the concentration error is considerably less than this, indicating that there was probably some internal cancellation of errors within the delay circuit.

Table D-1

ERRORS IN COMPUTED FEED CONCENTRATION AND STEAM FLOW RATE DUE TO INTEGRATOR GRID OFFSET VOLTAGES

Run No.	V_{co}	V_{cl}	Steady State Offset (volts)	Time (min.)	ΔC(% Sugar)	ΔS(lbs./min.)
6FF	5.040	4.877	-0.163	167	-0.0795	+.020
7FF	5.040	4.911	-0.129	205	-0.063	+.006
8FF	6.720	6.824	+0.104	151	+0.0506	-.008
9FF	6.800	6.829	+0.029	181	+0.014	-.007
10FF	3.800	3.777	-0.023	187	-0.011	+.002
11FFB	6.160	6.135	-0.025	145	-0.012	+.002

Fortunately the narrow calibration span on the feed refractometer reduces the significance of the feed concentration offset error. The approximate error in the computed steam flow rate, due to the feed concentration error, is also included in Table D-1.

2. Erroneous Settings for Controller Parameters

Prior to each experimental run, the product concentration set point, feed concentration, latent heat requirements and heat loss rate were set on potentiometers in the computer. These settings remained unchanged during the run. However, heat loss rates and the latent heat requirements of the feed both vary slightly under changing operating conditions and it has been shown that the delayed feed concentration can contain some error. It is therefore possible that some of these settings may be in error at the steady state conditions at the end of the run.

Attempts to correlate the observed product concentration error with a calculated error, based on the possible error in the steam flow rate, were not too successful. The reasons for this lies in the manner in which the change in steam flow requirements during a run must be calculated. The only way to estimate the change in the latent heat and heat loss steam requirements is to take the difference between the final and initial steady state calculated values of these parameters. However, the experimental errors which these values are subject to makes the small differences between them practically meaningless. The effect of the experimental errors can only be magnified when the individual steam change estimates are totalled and the net steam error is multiplied by a high sensitivity term (18.7% sugar change in product

concentration per pound of steam flow change as determined in run number 20L) to convert it to a possible concentration error.

Although the numbers in Table D-2 were reported to the third decimal point so that the differences could be obtained, they are useful only as an indication of how the observed steady state product concentration offsets may have resulted. There is a similarity between the calculated and observed concentration errors for some runs; for others there is not. Under no circumstances should the calculations in Table D-2 be considered accurate enough to use for adjusting experimental data.

Offset errors in feed concentration were avoided in runs 12FFB, 15FFB, and 16FFB by holding the initial condition voltage on the integrator once the integrator output reached the correct voltage.

Once two mode feedback control is added to the feedforward computer the product concentration must go to its set point if the control system is stable. This was the case in runs 12FFB and 15FFB; however, runs 11FFB and 16FFB did not reach steady state and did have some offset when the runs were stopped.

Table D-2

POSSIBLE ERRORS IN COMPUTED STEAM FLOWS

Run No.	Errors in Steam Flow Rate lbs./min.: Feed Concentration Offset	Latent Heat Setting Error	Heat Loss Setting Error	Net Steam Error	Possible Concentration Error % Sugar	Actual Product Concentration Error % Sugar
6FF	+0.020	+0.005	-0.018	+0.007	+0.13	-0.18
7FF	+0.006	-0.028	+0.022	0.0	0.0	+0.89
8FF	-0.008	+0.026	+0.013	+0.003	+0.06	0.0
9FF	-0.007	-0.025	-0.004	-0.036	-0.67	-0.40
10FF	+0.002	+0.005	+0.003	+0.010	+0.19	+0.19
11FFB	+0.002	-0.006	+0.007	+0.003	+0.06	*
12FFB	0.0	+0.010	+0.017	+0.027	+0.51	0.0
15FFB	0.0	-0.015	+0.011	-0.004	-0.07	0.0
16FFB	0.0	-0.006	-0.044	-0.050	-0.93	*

* Did not reach steady state.

APPENDIX E

EXPERIMENTAL DATA

A. Steady State and Transient Data and Controller Settings

The following experimental data for runs 10L to 18FB are sufficiently complete to allow the calculation of all the material and heat balances which follow in Appendix F. Although the transient data for the response of product concentration is included with the experimental data for every run, product flow, first effect bottoms flow, steam flow and first effect and separator liquid level transient data are not included. These data are presented in graphical form in the main body of this thesis. These data did not require any adjustment; it is therefore felt that they do not merit the space which would be required to reproduce them here in tabular form.

The data sheets for runs 11FFB to 18FB include an extra column of numbers entitled "adjusted product concentration." The reason for the adjustment has been explained in Appendix B and will not be repeated here. The calculated drift rate, on the basis of which the concentration adjustments were made, has been included along with the adjusted concentrations. The average drift rate for runs 11FFB to 18FB is -0.10% sugar per hour.

EXPERIMENTAL DATA FOR RUN NO. 10L

INITIAL CONDITIONS

C_o	2.95	W_1	1.965	T_o	94
W_o	3.00	O_2	.980	T_s	294
S	1.322	W_p	.983	T_{s1}	221
O_1	1.030	C_p	9.12	T_1	190
				T_2	157

Disturbance: Step Decrease in Condenser Pressure 4.41 to 3.00 Psia.

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Time (min.)	Product Conc. Weight Percent Sugar
0	9.12	40	10.11
4	9.48	50	10.29
8	9.59	60	10.53
12	9.72	75	10.62
16	9.76	90	10.63
20	9.80	106	10.67
25	9.90	110	10.71
30	10.00		

FINAL CONDITIONS

C_o	2.95	W_1	1.910	T_o	92
W_o	2.99	O_2	1.070	T_s	294
S	1.322	W_p	.865	T_{s1}	219
O_1	1.075	C_p	10.71	T_1	179
				T_2	141

CONTROLLER SETTINGS FOR RUN NO. 10L

Controller	Set Point*	% Prop. Band.	Reset (min.)	Other
FRC - 1	A	150	7/60	
FRC - 3	A	205	1/60	
FRC - 2	C	400	1/2	
LRC - 14	A	50	20	
LRC - 11	A	50	5	
LC - 15	A	30	10	
DVPRC - 21	A	75	4	
PRC - 22	A	10	1	
CRC - 2				O.L.
TC - 10	M			

PARAMETERS Second Effect Circulating Flow 268 lbs./min.
Condenser Cooling Water Flow 54.3 lbs./min.

*Automatic Control: A, Cascade Control: C, Manual Control: M

EXPERIMENTAL DATA FOR RUN NO. 20L

INITIAL CONDITIONS

C_o	3.27	W_1	2.098	T_o	94
W_o	3.07	O_2	.902	T_s	292
S	1.360	W_p	1.175	T_{s1}	220
O_1	.982	C_p	9.15	T_1	192
				T_2	160

Disturbance: Step Increase in Steam Flow Rate 1.37 to 1.563 lbs./min.

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Time (min.)	Product Conc. Weight Percent Sugar
0	9.15	70	11.53
5	9.16	80	11.70
10	9.45	90.5	11.90
16	9.70	105	12.12
20	9.88	120	12.22
25	10.06	135	12.36
30	10.28	150	12.54
40	10.61	165	12.60
52	11.07	180	12.75
60	11.20	195	12.76

FINAL CONDITIONS

C_o	3.25	W_1	1.97	T_o	92
W_o	3.070	O_2	1.095	T_s	292
S	1.563	W_p	.780	T_{s1}	221
O_1	1.130	C_p	12.76	T_1	192
				T_2	157

CONTROLLER SETTINGS FOR RUN NO. 20L

Controller	Set Point	% Prop. Band.	Reset (min.)	Other
FRC - 1	A	150	7/60	
FRC - 3	A	205	1/60	
FRC - 2	C	400	1/2	
LRC - 14	A	50	30	
LRC - 11	A	50	8	
LC - 15	A	30	10	
DVPRC - 21	A	75	4	
PRC - 22	A	10	1	
CRC - 2				O.L.
TC - 10	M			

PARAMETERS

Second Effect Circulation Flow = 282 lbs./min.
Condenser Cooling Water Flow = 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 30L

INITIAL CONDITIONS

C_o	4.32	W_1	1.925	T_o	93
W_o	2.90	O_2	.905	T_s	296
S	1.410	W_p	1.01	T_{s1}	219
O_1	1.02	C_p	11.96	T_1	191
				T_2	157.5

Disturbance: Step Decrease in Feed Concentration 4.32% to 3.06%

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Time (min.)	Product Conc. Weight Percent Sugar
0	12.16	70	10.30
6	12.15	80.5	10.05
10	12.12	90	9.88
15	12.01	105	9.70
20	11.87	120	9.55
25	11.75	135	9.35
31	11.65	150	9.25
40	11.35	165	9.10
50	10.88	180	9.10
60	10.65		

FINAL CONDITIONS

C_o	3.06	W_1	1.940	T_o	89
W_o	2.92	O_2	.930	T_s	296
S	1.410	W_p	1.01	T_{s1}	218.5
O_1	1.025	C_p	9.10	T_1	191
				T_2	157.5

CONTROLLER SETTINGS FOR RUN NO. 30L

Controller	Set Point	% Prop. Band.	Reset (min.)	Other
FRC - 1	A	150	7/60	
FRC - 3	A	200	1/60	
FRC - 2	C	400	1/2	
LRC - 14	A	50	20	
LRC - 11	A	25	5	
LC - 15	A	25	10	
DVPRC - 21	A	75	4	
PRC - 22	A	10	1	
CRC - 2				O.L.
TC - 10	M			

PARAMETERS
Circulating Flow = 282 lbs./min.
Cooling Water Flow = 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 40L

INITIAL CONDITIONS

C_o	2.90	W_1	1.940	T_o	94
W_o	2.92	0_2	.930	T_s	296
S	1.400	W_p	1.03	T_{s1}	218.5
0_1	1.015	C_p	9.00	T_1	191
				T_2	157

Disturbance: Step Increase in Feed Flow Rate 2.92 to 3.72 lbs./min.

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Time (min.)	Product Conc. Weight Percent Sugar
0	9.00	70	5.85
5	8.73	80.5	5.80
10.5	8.50	90	5.68
15	8.10	105	5.75
20.5	7.71	120.5	5.79
25	7.25	137	5.75
30	7.04	150	5.69
40	6.60	165	5.65
50	6.23	180	5.65
60	6.00		

FINAL CONDITIONS

C_o	2.95	W_1	2.85	T_o	92
W_o	3.72	0_2	.887	T_s	296
S	1.400	W_p	2.01	T_{s1}	217
0_1	.94	C_p	5.65	T_1	189
				T_2	157

CONTROLLER SETTINGS FOR RUN NO. 40L

Controller	Set Point	% Prop. Band.	Reset (min.)	Other
FRC - 1	A	150	7/60	
FRC - 3	A	200	1/60	
FRC - 2	C	400	1/2	
LRC - 14	A	50	20	
LRC - 11	A	25	5	
LC - 15	A	25	10	
DVPRC - 21	A	75	4	
PRC - 22	A	10	1	
CRC - 2				O.L.
TC - 10	M			

PARAMETERS Second Effect Circulating Flow = 282 lb./min.
Condenser Cooling Water Flow - 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 5FF

INITIAL CONDITIONS

C_o	3.07	W_1	2.030	T_o	93
W_o	2.98	O_2	.930	T_s	295.5
S	1.194	W_p	1.075	T_{s1}	221
O_1	.980	C_p	9.10	T_1	186
				T_2	156

Disturbance: Step Increase in Feed Concentration 3.07% to 4.31%

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Time (min.)	Product Conc. Weight Percent Sugar
0	9.10	60	8.05
4	9.12	71	7.96
8	9.05	80	7.91
12	8.97	90	7.90
16	8.95	105	7.90
20	8.90	120	7.92
25	8.75	135	7.90
30	8.60	150	7.92
40	8.40	156	7.92
50.5	8.16	180	7.92

FINAL CONDITIONS

C_o	4.31	W_1	2.34	T_o	95
W_o	2.98	O_2	.640	T_s	298
S	.840	W_p	1.700	T_{s1}	208
O_1	.665	C_p	7.92	T_1	179
				T_2	155

CONTROLLER SETTINGS FOR RUN NO. 5FF

Controller	Set Point	% Prop. Band.	Reset (min.)	Other
FRC - 1	C	150	7/60	
FRC - 3	A	205	1/60	
FRC - 2	C	400	1/2	
LRC - 14	A	50	20	
LRC - 11	A	100	4 (New Valve)	
LC - 15	A	60	10	
DVPRC - 21	A	75	4	
PRC - 22	A	10	1	
CRC - 2	C	125	.20 (Balanced)	
RC - 10	M			

PARAMETERS Second Effect Circulating Flow = 268 lbs./min.
Condenser Cooling Water Flow = 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 6FF

INITIAL CONDITIONS

C_o	3.10	W_1	2.20	T_o	92
W_o	2.99	O_2	.760	T_s	298
S	1.035	W_p	1.53	T_{s1}	211
O_1	.782	C_p	6.31	T_1	181
				T_2	155

Disturbance: Step Increase in Feed Composition 3.10% to 3.50%

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Time (min.)	Product Conc. Weight Percent Sugar
0	6.31	60	6.10
4	6.28	70	6.10
8	6.20	80	6.10
12	6.22	90	6.12
16	6.24	105	6.10
20	6.18	120	6.10
25	6.20	135	6.12
30	6.18	150	6.12
40	6.19	165	6.12
50	6.14		

FINAL CONDITIONS

C_o	3.50	W_1	2.32	T_o	93
W_o	2.99	O_2	.630	T_s	298
S	0.921	W_p	1.76	T_{s1}	209
O_1	.565	C_p	6.15	T_1	180
				T_2	155

CONTROLLER SETTINGS FOR RUN NO. 6FF

Controller	Set Point	% Prop. Band.	Reset (min.)	Other
FRC - 1	C	150	7/60	
FRC - 3	A	205	1/60	
FRC - 2	C	400	1/2	
LRC - 14	A	50	20	
LRC - 11	A	100	5	
LC - 15	A	30	10	
DVPRC - 21	A	75	4	
PRC - 22	A	10	1	
CRC - 2	C	100	.2(Balanced)	
TC - 10	M			

PARAMETERS
Second Effect Circulating Flow = 268 lbs./min.
Condenser Cooling Water Flow = 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 7FF

INITIAL CONDITIONS

C_o	3.50	W_1	2.375	T_o	93
W_o	2.99	O_2	.603	T_s	297
S	0.915	W_p	1.750	T_{s1}	208
O_1	.672	C_p	6.19	T_1	180
				T_2	156

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Time (min.)	Product Conc. Weight Percent Sugar
0	6.19	70	9.75
4	6.30	80	9.95
8	6.52	90	10.04
12	6.79	105	10.34
16	7.00	120	10.51
20	7.26	135.5	10.65
25	7.65	150	10.75
30	7.91	168	10.88
40	8.59	180.5	10.90
50	9.05	205	10.89
60	9.40		

FINAL CONDITIONS

C_o	3.45	W_1	1.985	T_o	93
W_o	3.00	O_2	1.00	T_s	294
S	1.316	W_p	.990	T_{s1}	220.5
O_1	1.060	C_p	10.90	T_1	188
				T_2	156

CONTROLLER SETTINGS FOR RUN NO. 7FF

Controller	Set Point	% Prop. Band.	Reset (min.)	Other
FRC - 1	C	150	7/60	
FRC - 3	A	205	1/60	
FRC - 2	C	400	1/2	
LRC - 14	A	50	20	
LRC - 11	A	50	5	
LC - 15	A	30	10	
DVPRC - 21	A	75	4	
PRC - 22	A	10	1	
CRC - 2	C	100	.2(Balanced)	
TC - 10	M			

PARAMETERS Second Effect Circulating Flow = 268 lbs./min.
Condenser Cooling Water Flow 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 8FF

INITIAL CONDITIONS

C_o	3.45	W_1	1.98	T_o	94
W_o	2.980	0_2	1.000	T_s	294
S	1.315	W_p	1.001	T_{s1}	221
0_1	1.06	C_p	10.98	T_1	189
				T_2	157

Disturbance: Step Increase in Feed Composition 3.45% to 4.25%

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Time (min.)	Product Conc. Weight Percent Sugar
0	10.90	60	10.82
4	10.98	70	10.80
8	11.04	80	10.85
12	11.04	90	10.81
16	11.06	105	10.80
20	11.06	120	10.79
25	11.10	135	10.70
30	11.11	150	10.70
40	11.07	165	10.70
50	10.99		

FINAL CONDITIONS

C_o	4.25	W_1	2.10	T_o	97
W_o	2.98	0_2	.870	T_s	296
S	1.149	W_p	1.280	T_{s1}	216
0_1	.916	C_p	10.70	T_1	184
				T_2	156

CONTROLLER SETTINGS FOR RUN NO. 8FF

Controller	Set Point	% Prop. Band.	Reset (min.)
FRC - 1	C	150	7/60
FRC - 3	A	205	1/60
FRC - 2	C	400	1/2
LRC - 14	A	50	20
LRC - 11	A	50	5
LC - 15	A	30	10
DVPRC - 21	A	75	4
PRC - 22	A	10	1
CRC - 2	C	100	.2 (Balanced)
TC - 10	M		

PARAMETERS Second Effect Circulating Flow (72%) 268 lbs./min.
Condenser Cooling Water Flow 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN No. 9FF

INITIAL CONDITIONS

C_o	4.30	W_1	2.085	T_o	97
W_o	2.98	O_2	.870	T_s	296
S	1.149	W_p	1.26	T_{s1}	216
O_1	.910	C_p	10.70	T_1	184
				T_2	156

Disturbance: Step Increase in Feed Flow Rate 2.98 to 3.85 lbs./min.

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Time (min.)	Product Conc. Weight Percent Sugar
0	10.66	60	10.05
4	10.75	70	10.00
8	10.80	80	10.01
12	10.88	90	10.08
16	10.90	105	10.20
20	10.87	120	10.26
25	10.84	135	10.29
30.5	10.67	150	10.31
40	10.45	165	10.32
50	10.20	180	10.30

FINAL CONDITIONS

C_o	4.30	W_1	2.73	T_o	96
W_o	3.85	O_2	1.110	T_s	291
S	1.470	W_p	1.665	T_{s1}	222.5
O_1	1.120	C_p	10.30	T_1	189
				T_2	155

CONTROLLER SETTINGS FOR RUN NO. 9FF

Controller	Set Point	% Prop. Band.	Reset (min.)
FRC - 1	C	150	7/60
FRC - 3	A	205	1/60
FRC - 2	C	400	1/2
LRC - 14	A	50	20
LRC - 11	A	50	5
LC - 15	A	30	10
DVPRC - 21	A	75	4
PRC - 22	A	10	1
CRC - 2	C	100	.20 (Balanced)
TC - 10	M		

PARAMETERS Second Effect Circulating Flow = 268 lbs./min.
Condenser Cooling Water Flow = 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 10FF

INITIAL CONDITIONS

C_o	2.70	W_1	1.890	T_o	96
W_o	2.98	O_2	1.070	T_s	293
S	1.388	W_p	.875	T_{s1}	221
O_1	1.12	C_p	10.42	T_1	190
				T_2	156

Disturbance: Step Decrease in Product Concentration Set Point 10.50% to 8.50%

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Time (min.)	Product Conc. Weight Percent Sugar
0	10.42	71	8.99
4	10.33	80	8.91
8	10.23	90	8.90
12	10.18	105	8.79
16	10.05	120	8.65
20	9.98	135.5	8.70
25	9.81	150	8.69
30	9.70	165	8.75
40	9.45	180	8.69
50	9.25	187	8.69
62	9.01		

FINAL CONDITIONS

C_o	2.78	W_1	1.95	T_o	95
W_o	2.98	O_2	.985	T_s	294
S	1.292	W_p	.970	T_{s1}	220
O_1	1.05	C_p	8.69	T_1	187
				T_2	156

CONTROLLER SETTINGS FOR RUN NO. 10FF

Controller	Set Point	% Prop. Band.	Reset (min.)
FRC - 1	C	150	7/60
FRC - 3	A	205	1/60
FRC - 2	C	400	1/2
LRC - 14	A	50	20
LRC - 11	A	50	5
LC - 15	A	30	10
DVPRC - 21	A	75	4
PRC - 22	A	10	1
CRC - 2	C	100	.20 (Balanced)
TC - 10	M		

PARAMETERS
Second Effect Circulating Flow = 268 lbs./min.
Condenser Cooling Water Flow = 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 11FFB

INITIAL CONDITIONS

C_o	3.44	W_1	2.00	T_o	96
W_o	2.97	O_2	.958	T_s	294
S	1.260	W_p	1.100	T_{s1}	216
O_1	1.000	C_p	10.14	T_1	186
				T_2	156

Disturbance: Step Increase in Feed Concentration 3.44 to 3.90%

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Conc.	Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Conc.
0	10.05	10.05	50	10.08	10.06
4	10.11	10.11	60	10.10	10.08
8	10.14	10.14	73	10.13	10.11
12	10.14	10.14	80	10.19	10.16
16	10.14	10.13	95	10.20	10.17
20	10.10	10.09	105	10.30	10.26
25	10.15	10.14	120	10.20	10.16
30	10.12	10.11	136	10.25	10.20
40	10.00	9.98	145	10.30	10.25

Refractometer Drift Rate = -.0223% Sugar/Hour

FINAL CONDITIONS

C_o	3.95	W_1	2.05	T_o	94
W_o	2.97	O_2	.920	T_s	296
S	1.219	W_p	1.155	T_{s1}	219
O_1	.970	C_p	10.30	T_1	186
				T_2	156

CONTROLLER SETTINGS FOR RUN NO. 11FFB

Controller	Set Point	% Prop. Band.	Reset (min.)
FRC - 1	C	150	7/60
FRC - 3	A	205	1/60
FRC - 2	C	400	1/2
LRC - 14	A	50	20
LRC - 11	A	50	5
LC - 15	A	30	10
DVPRC - 21	A	75	4
PRC - 22	A	10	1
CRC - 2	C	50	5
TC - 10	M		

PARAMETERS
Second Effect Circulating Flow = 268 lbs./min.
Condenser Cooling Water Flow - 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 12FFB

INITIAL CONDITIONS

C_o	3.38	W_1	2.05	T_o	91
W_o	2.97	O_2	.888	T_s	295.5
S	1.194	W_p	1.170	T_{s1}	221
O_1	.945	C_p	9.02	T_1	187
				T_2	157

Disturbance: Step Increase in Feed Concentration 3.38% to 3.80%

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Conc.	Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Conc.
0	9.02	9.02	60	9.14	9.12
4	9.06	9.06	70	9.13	9.11
8	8.98	8.98	80	9.12	9.09
12	8.95	8.95	90	9.11	9.08
16	8.95	8.94	100	9.20	9.16
20	8.95	8.94	110	9.10	9.06
25.5	8.89	8.88	120	9.11	9.07
30	8.89	8.88	137	9.15	9.11
40	8.95	8.94	141	9.15	9.10
50	9.00	8.98			

Refractometer Drift Rate = -.0216% Sugar/Hour

FINAL CONDITIONS

C_o	3.80	W_1	2.090	T_o	94
W_o	2.97	O_2	.850	T_s	297
S	1.125	W_p	1.245	T_{s1}	219
O_1	.898	C_p	9.15	T_1	187
				T_2	157

CONTROLLER SETTINGS FOR RUN NO. 12FFB

Controller		Set Point	% Prop. Band.	Reset (min.)
FRC	- 1	C	150	7/60
FRC	- 3	A	205	1/60
FRC	- 2	C	400	1/2
LRC	- 14	A	50	12
LRC	- 11	A	150	8 (Valve Cleaned)
LC	- 15	A	30	10
DVPRC	- 21	A	75	4
PRC	- 22	A	10	1
CRC	- 2	C	100	10
TC	- 10	M		

PARAMETERS
Second Effect Circulating Flow = 268 lbs./min.
Condenser Cooling Water Flow = 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 13FFB

INITIAL CONDITIONS

C_o	3.80	W_1	2.100	T_o	94
W_o	2.97	O_2	.860	T_s	297
S	1.140	W_p	1.250	T_{s1}	220
O_1	.900	C_p	8.89	T_1	188
				T_2	158

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Compositions
0	9.25	9.10
4	9.40	9.25
8	9.50	9.35
12	9.58	9.43
16.5	9.46	9.30
20	9.45	9.29
25.5	9.23	9.07
30	9.10	8.94
36	8.99	
40	8.89	8.73
50	8.90	8.73
60	8.98	8.81

Set Point = 9.10

.15% error - drift rate of -.0216%/hr.

Stop: System Unstable

CONTROLLER SETTINGS FOR RUN NO. 13FFB

Controller	Set Point	% Prop. Band.	Reset (min.)
FRC - 1	C	150	7/60
FRC - 3	A	205	1/60
FRC - 2	C	400	1/2
LRC - 14	A	50	12
LRC - 11	A	150	8
LC - 15	A	30	10
DVPRC - 21	A	75	4
PRC - 22	A	10	1
CRC - 2	C	130	10
TC - 10	M		

PARAMETERS

Second Effect Circulating Flow = 268 lbs./min.

Condenser Cooling Water FLow = 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 14FFB

INITIAL CONDITIONS

C_o	3.70	W_1	2.690	T_o	98
W_o	3.73	O_2	1.135	T_s	292
S	1.455	W_p	1.650	T_{s1}	227
O_1	1.140	C_p	8.98	T_1	190
				T_2	157

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar
0	8.98
4	8.91
8	8.86
12	8.90
16	8.99
20	9.13
28	9.57
35	9.90
41	9.90

Drift Corrections (-.0216%/hr.)
Negligible over 40 mins.
(.01% Sugar)

Stop: System Unstable

CONTROLLER SETTINGS FOR RUN NO. 14FFB

Controller	Set Point	% Prop. Band.	Reset (min.)
FRC - 1	C	150	7/60
FRC - 3	A	205	1/60
FRC - 2	C	400	1/2
LRC - 14	A	50	12
LRC - 11	A	150	8
LC - 15	A	30	10
DVPRC - 21	A	75	4
PRC - 22	A	10	1
CRC - 2	C	400	2.0
TC - 10	M		

PARAMETERS
Second Effect Circulating Flow = 268 lbs./min.
Condenser Cooling Water Flow = 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 15FFB

INITIAL CONDITIONS

C_o	4.31	W_1	2.16	T_o	100
W_o	3.00	0_2	.750	T_s	298
S	1.001	W_p	1.260	T_{s1}	212
0_1	.799	C_p	9.10	T_1	185
				T_2	157

Disturbance: Step Increase in Feed Flow 3.00 to 3.870 lbs./min.

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Conc.	Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Conc.
0	9.10	9.10	60	9.13	8.95
4	9.20	9.19	70	9.10	8.89
8	9.32	9.30	80	9.21	8.97
12	9.38	9.34	90	9.40	9.23
16	9.44	9.39	100	9.55	9.25
20	9.41	9.35	110	9.50	9.17
25	9.40	9.32	120	9.52	9.16
30	9.35	9.26	135	9.49	9.08
40	9.22	9.10	150	9.56	9.11
50	9.13	8.98	165	9.60	9.10

Refractometer Drift Rate = -.18%/hr.

FINAL CONDITIONS

C_o	4.26	W_1	2.81	T_o	100
W_o	3.870	0_2	1.050	T_s	294
S	1.360	W_p	1.80	T_{s1}	222
0_1	1.060	C_p	9.60	T_1	189
				T_2	157

CONTROLLER SETTINGS FOR RUN NO. 15FFB

Controller		Set Point	% Prop. Band.	Reset (min.)
FRC	- 1	C	150	7/60
FRC	- 3	A	205	1/60
FRC	- 2	C	400	1/2
LRC	- 14	A	50	12
LRC	- 11	A	120	5
LC	- 15	A	30	10
DVPRC	- 21	A	75	4
PRC	- 22	A	10	1
CRC	- 2	C	400	30
TC	- 10	M		

PARAMETERS Second Effect Circulating Flow = 268 lbs./min.
Condenser Cooling Water Flow = 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 16FFB

INITIAL CONDITIONS

C_o	2.73	W_1	2.000	T_o	96
W_o	2.97	O_2	1.030	T_s	295
S	1.250	W_p	.970	T_{s1}	240
O_1	1.010	C_p	8.20	T_1	185
				T_2	154

Disturbance: Step Increase in Product Composition Set Point 8.35 to 12.08%

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Conc.	Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Conc.
0	8.20	8.20	80	13.30	13.17
5	8.34	8.33	90	13.50	13.35
8	8.45	8.44	105	13.61	13.43
12	8.62	8.60	121	13.30	13.10
16	8.84	8.82	135	13.19	12.97
20	9.02	9.29	150	12.93	12.68
25	9.33		165	12.79	12.51
30	9.69	9.64	180	12.63	12.33
40.5	10.36	10.29	195.5	12.55	12.22
50	11.17	11.09	210	12.35	12.00
60	11.70	11.60	225	12.29	11.91
70	12.57	12.45	235	12.30	11.91

Refactometer Drift Rate = -.10%/hr.

FINAL CONDITIONS

C_o	2.73	W_1	1.850	T_o	97.5
W_o	2.97	O_2	1.170	T_s	292.5
S	1.440	W_p	.702	T_{s1}	237
O_1	1.150	C_p	12.30	T_1	189
				T_2	154

CONTROLLER SETTINGS FOR RUN NO. 16FFB

Controller	Set Point	% Prop. Band.	Reset (min.)
FRC - 1	C	150	7/60
FRC - 3	A	205	1/60
FRC - 2	C	400	1/2
LRC - 14	A	50	12
LRC - 11	A	120	5
LC - 15	A	30	10
DVPRC - 21	A	75	4
PRC - 22	A	10	1
CRC - 2	C	400	30
TC - 10	M		

PARAMETERS
Second Effect Circulating Flow (72%) 268 lbs./min.
Condenser Cooling Water Flow 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 17FB

INITIAL CONDITIONS

C_o	4.53	W_1	2.180	T_o	97
W_o	2.99	0_2	.815	T_s	298
S	1.050	W_p	1.375	T_{s1}	219
0_1	.800	C_p	9.71	T_1	182
				T_2	155

Disturbance: Step Decrease in Feed Concentration 4.53% to 3.08%

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Conc.	Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Conc.
0	9.71	9.71	70	10.06	9.90
4	9.70	9.69	80	9.98	9.80
8	9.64	9.62	90	9.87	9.66
12	9.54	9.51	105	9.90	9.66
16	9.43	9.40	120	10.15	9.88
20	9.33	9.29	135	10.20	9.90
25	9.30	9.24	150	10.18	9.84
30	9.38	9.31	164	10.11	9.75
40	9.46	9.37	180	10.17	9.76
50	9.86	9.74	189	10.19	9.76
60	10.08	9.94			

Refractometer Drift Rate = -.137%/hr.

FINAL CONDITIONS

C_o	3.08	W_1	1.955	T_o	97
W_o	3.00	0_2	1.070	T_s	294
S	1.357	W_p	.870	T_{s1}	226
0_1	1.090	C_p	10.19	T_1	188
				T_2	156

CONTROLLER SETTINGS FOR RUN NO. 17FB

Controller	Set Point	% Prop. Band.	Reset (min.)	Rate (min.)
FRC - 1	C	390	1/6	
FRC - 3	A	205	1/60	
FRC - 2	C	400	1/2	
LRC - 14	A	50	12	
LRC - 11	A	120	5	
LC - 15	A	30	10	
DVPRC - 21	A	75	4	
PRC - 22	A	10	1	
CRC - 2	C	130	7	5
TC - 10	M			

PARAMETERS Second Effect Circulating Flow (72%) 268 lbs./min.
Condenser Cooling Water Flow 54.3 lbs./min.

EXPERIMENTAL DATA FOR RUN NO. 18FB

INITIAL CONDITIONS

C_o	3.08	W_1	1.900	T_o	97
W_o	3.00	0_2	1.085	T_s	294
S	1.355	W_p	.840	T_{s1}	226
0_1	1.090	C_p	10.35	T_1	187
				T_2	156

Disturbance: Step Increase in Feed Flow

TRANSIENT DATA

Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Conc.	Time (min.)	Product Conc. Weight Percent Sugar	Adjusted Product Conc.
0	10.35	9.78	71	10.56	9.83
5	10.29	9.71	81	11.00	10.25
8	10.19	9.60	92	11.01	10.23
12	11.00	9.41	105	10.68	9.88
16	9.89	9.29	120	10.26	9.42
20	9.70	9.09	135	10.27	9.39
25	9.51	8.89	150	10.50	9.59
30	9.40	8.76	165	10.78	9.83
40	9.32	8.66	180	10.81	9.83
50	9.60	8.91	195	10.62	9.60
60	10.04	9.31	305	10.60	9.56

Drift Rate = -.137%/hr. + 0.57% Sugar Set Point Error @Start

FINAL CONDITIONS

C_o	3.00	W_1	2.250	T_o	95
W_o	3.48	0_2	1.230	T_s	291
S	1.610	W_p	1.055	T_{s1}	229
0_1	1.195	C_p	10.60	T_1	191
				T_2	156

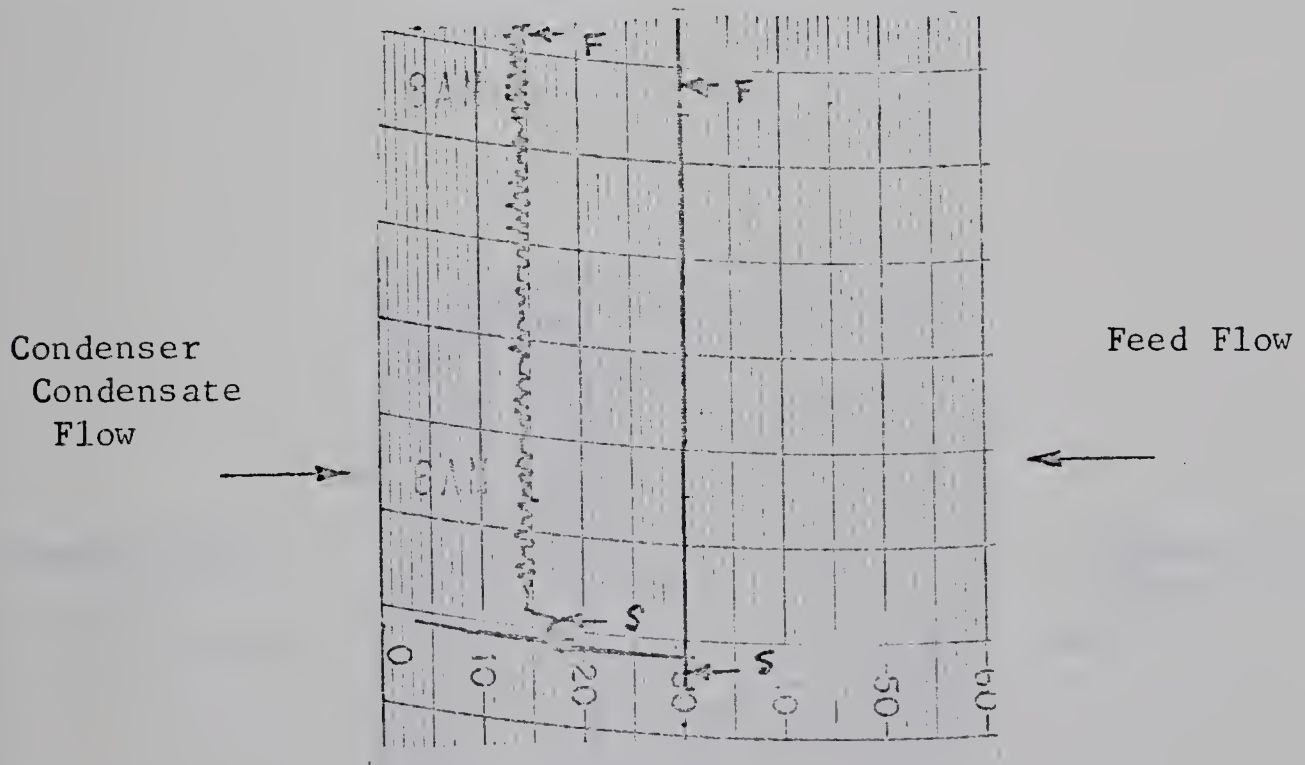
CONTROLLER SETTINGS FOR RUN NO. 18FB

Controller		Set Point	% Prop. Band.	Reset (min.)	Rate (min.)
FRC	- 1	C	390	1/6	
FRC	- 3	A	205	1/60	
FRC	- 2	C	400	1/2	
LRC	- 14	A	50	8	
LC	- 11	A	120	5	
LC	- 15	A	30	10	
DVPRC	- 21	A	75	4	
PRC	- 22	A	10	1	
CRC	- 2	C	130	7	5
TC	- 10	M			

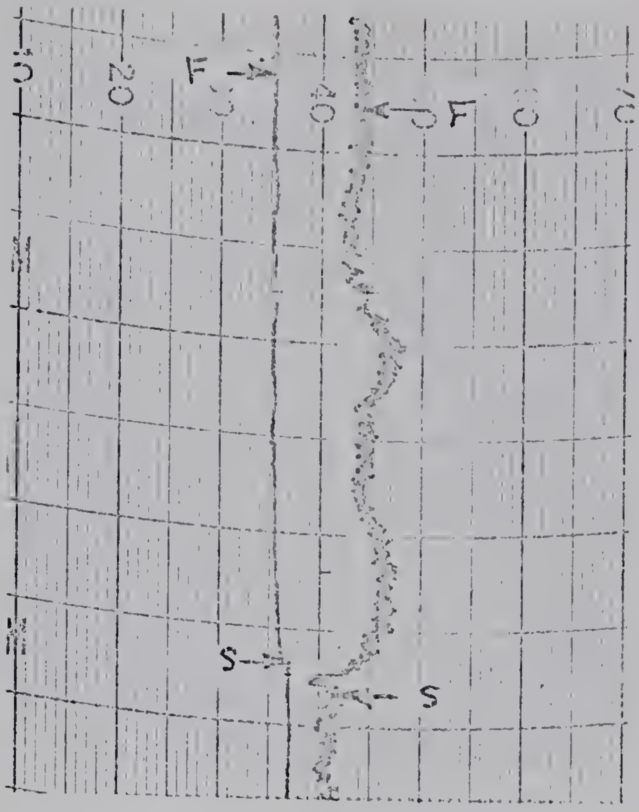
PARAMETERS
Second Effect Circulating Flow = 268 lbs./min.
Condenser Cooling Water FLow = 54.3 lbs./min.

B. Typical Recordings of Process Variables

The first five charts below are taken from experimental run number 10FF. The start and finish times are marked on each record with an "S" and an "F" respectively. The variables represented by the recordings are indicated to the right for the upper curve and to the left for the lower curve. Where there was some oscillation in the signal (as in the first effect overhead and product flows) the average pen position was read as the correct value for the variable. The time scale as indicated on the charts is 3/4 of an inch per hour. Chart percent readings were converted to actual values using the calibration curves of Appendix B. The offset in starting times between the two curves on each chart is the result of the recording pen offset.

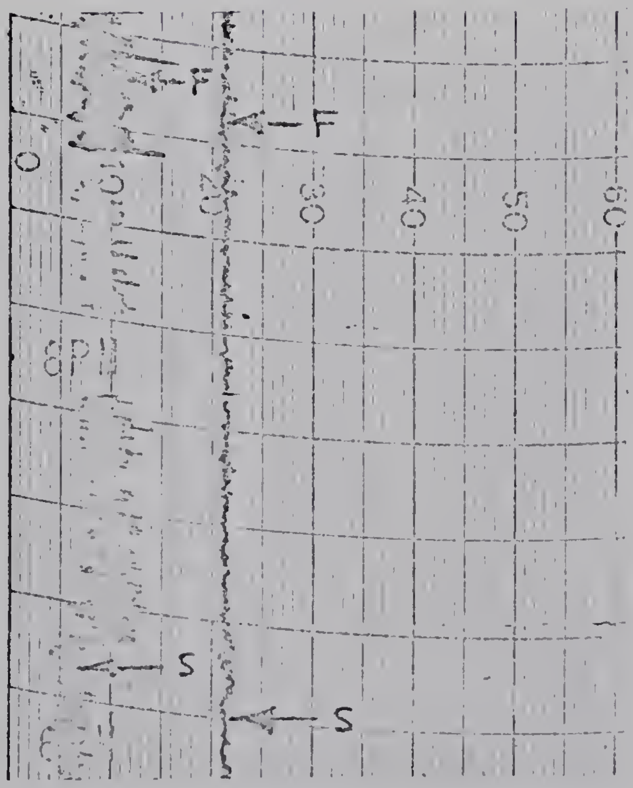


Comment: The disturbance in the condensate flow rate prior to Run 10FF resulted when the condenser condensate pump was turned off. This pump was used for only a short period between experimental runs to transfer condensate from the previous run to the large water tank. The steady state flow rate, previous to operation of the pump, and the condensate flow at the start of Run No. 10FF were identical.



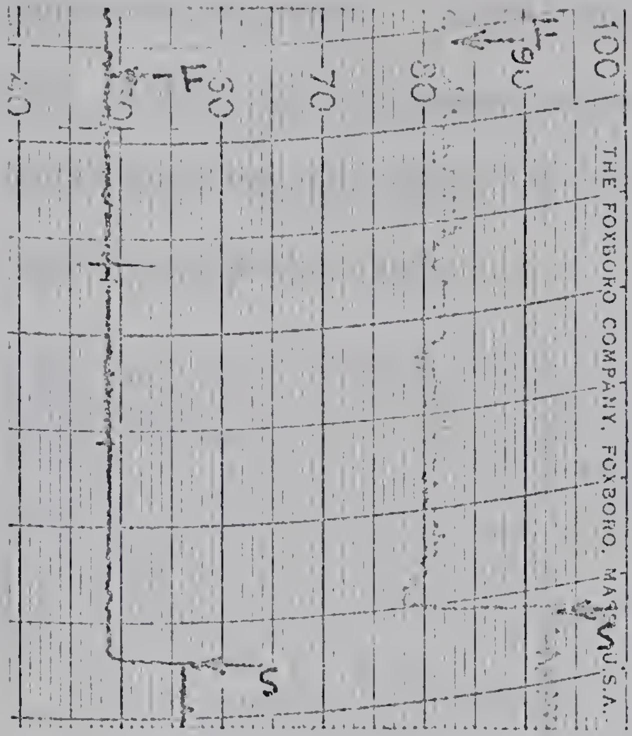
First Effect
Absolute
Pressure

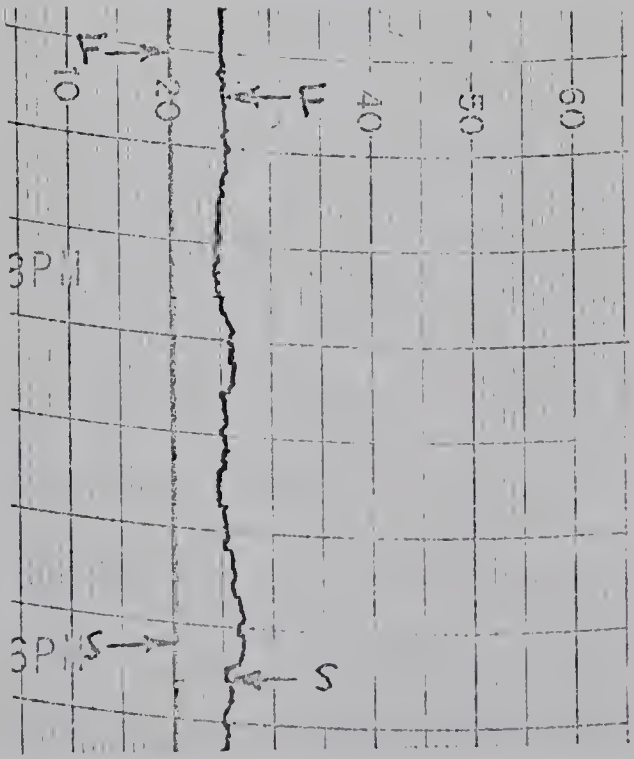
First Effect
Bottoms Flow



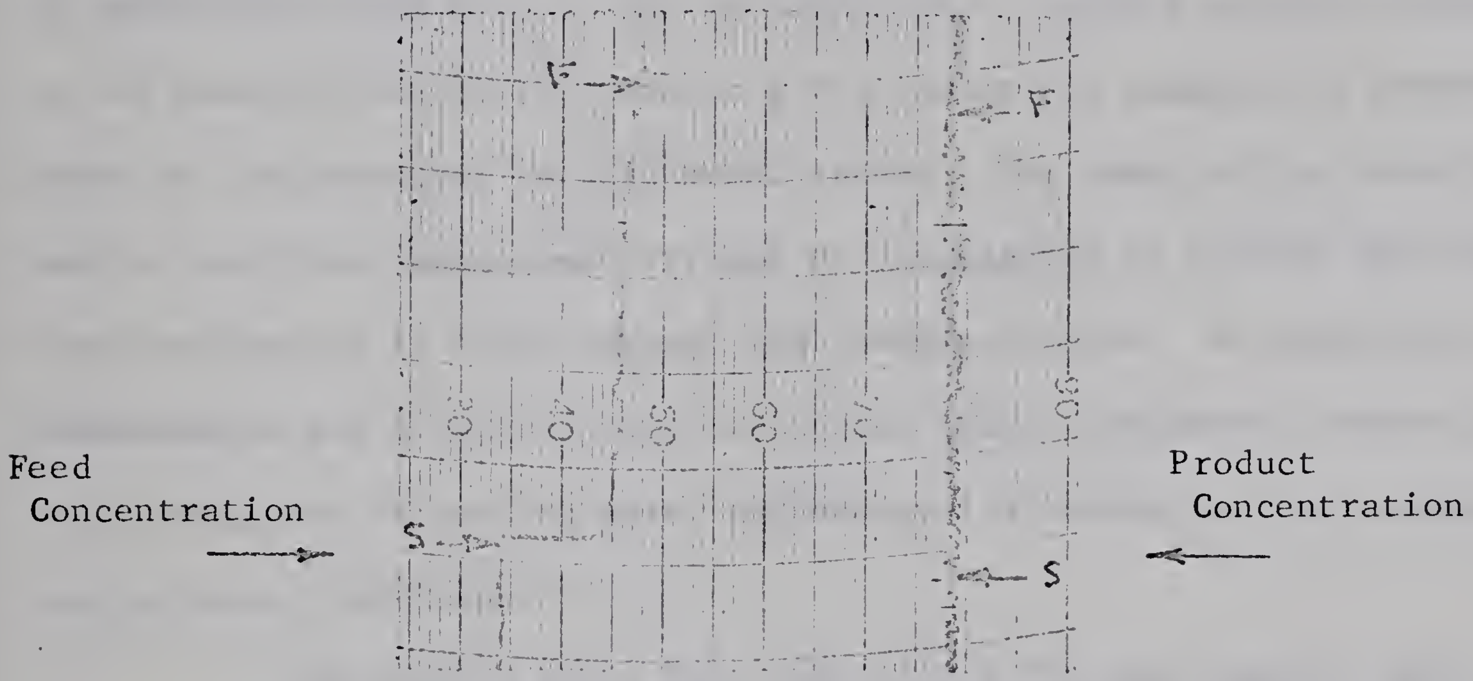
Product Flow

Separator Liquid
Level


Steam Flow
First Effect Overhead Flow (Condensate)
100
90
80
70
60
50
40
F
S
THE FOXBORO COMPANY, FOXBORO, MASS. U.S.A.


Condenser Absolute Pressure
First Effect Liquid Level
10
20
40
50
60
F
S
SP
SP

Unfortunately the product and feed concentration recordings from Run No. 10FF were discarded by error. Therefore the concentration recordings from Run No. 11FFB are presented below. The drift in the feed concentration recording was due to a 2°F decrease in feed temperature during the experimental run.



APPENDIX F

CALCULATIONS

A. Steady State Calculations

The evaporator is equipped with sufficient flow meters to provide all the information required for the calculation of balances around each effect and around the evaporating unit. A heat balance can be made around each effect, but an independent, overall enthalpy balance is not possible due to the lack of a flow meter and temperature measurements on the condenser cooling water stream. The same cooling water is used to cool both condensate streams by circulating it through the rundown tank before it flows through the condenser tubes. No temperature measurements are available from the second effect condensate stream; the only indication of cooling water temperature is located on the condenser cooling water discharge.

The cooling water flow rate with a 90% open control valve was determined to be 54.3 pounds per minute when the evaporator was not operating. The water temperature was 66°F. Even if these measurements were considered sufficiently accurate for an overall enthalpy balance, assumptions regarding the temperature of the second effect condensate would be required. In view of these difficulties, the overall heat balance does not include the enthalpies of cooling water streams. The feedforward composition controller does not require any information regarding the condenser system; therefore the above ommission does not affect the calculations made to check out the accuracy of the computing system and its steady state model.

The following calculations are from experimental Run No. 12FFB.

Sample Calculations: Experimental Run No. 12FFB

Basis for Calculations: 1 Minute

1. Material Balances

(i) Overall Balance

(a) Initial Steady State

Feed		2.97 lbs./min.
Product	1.170	
O_1	0.945	
O_2	0.888	
Total Products		3.003 lbs./min.

Error : + 0.033 lbs./min. (1.10%)

(b) Final Steady State

Feed		2.97 lbs./min.
Product	1.245	
O_1	0.898	
O_2	0.850	
Total Products		2.993 lbs./min.

Error : + 0.023 lbs./min. (.785%)

(ii) Sugar Balance

(a) Initial Steady State

Input	2.97 x 0.0338 =	0.1003 lbs./min.
Output	1.170 x 0.0902 =	0.1056 lbs./min.

Error : + 0.0053 lbs./min. (5.30%)

(b) Final Steady State

Input 2.97 x 0.0380 = 0.1129

Output 1.245 x 0.0915 = 0.1140

Error : + 0.0011 (.98%)

(iii) First Effect Balance

(a) Initial Steady State

Feed 2.97 lbs./min.

W_1 2.05

O_1 0.945

Total Output 2.995 lbs./min.

Error : + 0.025 lbs./min. (.84%)

(b) Final Steady State

Feed 2.97 lbs./min.

W_1 2.090

O_1 0.898

Total Output 2.988 lbs./min.

Error : + 0.018 lbs./min. (.61%)

(iv) Second Effect Balance

(a) Initial Steady State

W_1 2.050 lbs./min.

Product 1.170

O_2 .888

Total output 2.058 lbs./min.

Error : + 0.008 (.39%)

(b) Final Steady State

W_1 2.090 lbs./min.

Product 1.245

O_2 .850

Total Output 2.095 lbs./min.

Error : + 0.005 (.24%)

2. Heat Balances

(i) First Effect

(a) Initial Steady State

Enthalpy of supply steam (296°F) = 1178.6 Btu/lb.

Enthalpy of steam condensate (221°F) = 189.1 Btu/lb.

Heat supply per pound of steam 989.5 Btu/lb.

Heat Input = 1.1946 x 989.5 = 1180 Btu/min.

Latent Heat to Feed = 2.97 (187 - 91) = 285

Heat Consumed by Boiling 932

Total Heat Consumed 1217 Btu/min.

-37 Btu/min.

Heat Transfer Coefficient (Based on Total Heat Consumed)

$$U_1 = \frac{1180 \text{ Btu/min.} \times 60 \text{ min./hr.}}{9.88 \text{ ft.}^2 \times (221^\circ - 187^\circ)}$$

$$= 211 \text{ Btu/hr. ft.}^2 \,^\circ\text{F}$$

(b) Final Steady State

Enthalpy of supply steam (297°F) = 1178.8 Btu/lb.

Enthalpy of steam condensate (219°F) = 187.1 Btu/lb.

Heat supply per pound of steam = 991.7 Btu/lb.

Heat input 1.125 x 991.7 = 1118 Btu/min.

Latent Heat to Feed = 2.97 (187 - 94) = 276

Heat Consumed by Boiling 0.898 x 986 = 886

Total Heat Consumed 1162 Btu/min.

Heat Losses at First Effect -44Btu/min.

Heat Transfer Coefficient (Based on Total Heat Consumed)

$$U_1 = \frac{1118 \text{ Btu/min.} \times 60 \text{ min./hr.}}{9.88 \text{ ft.}^2 \times (219^\circ - 187^\circ)}$$

$$U_1 = 212 \text{ Btu/hr. ft.}^{2\circ}\text{F}$$

(ii) Second Effect

(a) Initial Steady State

Condensation Heat Input 932

Latent Heat Input 29

Total Heat Input 961 Btu/min.

Heat Consumed by Boiling 892 Btu/min.

Heat Losses at Second Effect 69 Btu/min.

Inter-Stage Heat Losses = 2.05 (187 - 171) = 33 Btu/min.

Heat Transfer Coefficient (Based on Condensation Heat Input)

$$U_2 = \frac{932 \text{ Btu/min.} \times 60 \text{ min./hr.}}{4.10 \text{ ft.}^2 \times (187^\circ - 157^\circ)}$$

$$= 455 \text{ Btu/hr. ft.}^{2\circ}\text{F}$$

(b) Final Steady State

Condensation Heat Input 886

Latent Heat Input 31

Total Heat Input 917 Btu/min.

Heat Consumed by Boiling 855 Btu/min.

Heat Losses at Second Effect 62 Btu/min.

Inter-Stage Heat Losses = 2.09 (187 - 172) = 31 Btu/min.

Heat Transfer Coefficient (Based on Condensation Heat Input)

$$U_2 = \frac{886 \text{ Btu/min.} \times 60 \text{ min./hr.}}{4.10 \text{ ft.}^2 \times (187° - 157°)}$$

$$= 432 \text{ Btu/hr. ft.}^2 °F$$

(iii) Overall

Enthalpies Relative to Water at 32°F (0.0)

Comment: There is insufficient experimental data available about the condensate and cooling water streams to allow this heat balance to be truly an "overall" balance. Therefore, cooling water is not brought into the balance and the first effect overhead (condensate) and the second effect overhead (vapor) streams are considered to exchange no heat with the cooling water.

(a) Initial Steady State

Inputs:	Net steam heat input	1180	
	Feed heat input 2.97 x 58.99 =	175	
	Total Heat Input		1355 Btu/min.

Assumption: Since the temperature of the condensate in the second effect steam chest is unknown, assume its temperature to be equal to that of the condensing steam.

Outputs:

First Effect Overhead (Condensate)	.945 x 154.94 =	146
Second Effect Overhead	.888 x 1129.0 =	1001
Product	1.170 x 124.89 =	146
Heat Losses		65
Total Heat Output	1358 Btu/min.	
Balance Error	+3 Btu/min.	

(b) Final Steady State

Inputs:	Net steam heat input	1118	
	Feed heat input 2.97 x 61.98 =	184	
	Total Heat Input		1302 Btu/min.

Outputs:		
First Effect Overhead (Condensate)	.898 x 154.94 =	139
Second Effect Overhead	.850 x 1129.0 =	960
Product	1.245 x 124.89 =	155
Heat Losses		49
Total Heat Output		1303 Btu/min.
Balance Error		+1 Btu/min.

3. Calculated Steam Requirements

(a) Initial Steady State

(i) Steam to Concentrate Feed to Set Point Concentration

$$S_{MB} = W_o K_s (1-C_o/C_{psp})$$

$$= 2.97 \times .50 \left(1 - \frac{.0338}{.0902}\right) = 0.927 \text{ lbs./min.}$$

(ii) Steam to Supply Feed Latent Heat Requirements

$$S_{LH} = \frac{285 \text{ Btu/min.}}{989.5 \text{ Btu/lb.}} = 0.288 \text{ lbs./min.}$$

(iii) $$S_{HL} = \frac{65 \text{ Btu/min.}}{989.5 \text{ Btu/lb.}} = 0.066 \text{ lbs./min.}$$

Total Calculated Steam Requirements 1.281 lbs./min.

Measured

Steam Flow 1.194 lbs./min.

Percent Error 7.27%
(Based on Measured Steam Flow)

(b) Final Steady State

(i) Steam to Concentrate Feed to Set Point Concentration

$$S_{MB} = W_o K_s (1 - C_o / C_{psp})$$

$$= 2.97 \times .50 \left(1 - \frac{.0380}{.0915}\right) = 0.868 \text{ lbs./min.}$$

(ii) Steam to Supply Feed Latent Heat Requirements

$$S_{LH} = \frac{276 \text{ Btu/min.}}{991.7 \text{ Btu/lb.}} = 0.278 \text{ lbs./min.}$$

(iii) Steam to Offset Heat Losses

$$S_{HL} = \frac{49 \text{ Btu/min.}}{991.7 \text{ Btu/lb.}} = 0.049 \text{ lbs./min.}$$

Total Calculated Steam Requirements	1.195 lbs./min.
Measured Steam Flow	1.125 lbs./min.
Percent Error (Based on Measured Steam Flow)	6.23%

The following section summarizes the above calculations as they were made for each experimental run .

4. Summary of Steady State Calculations

Run No. 10L

	START		FINISH	
A. Material Balances (lbs./min.)				
1. Overall				
Feed	3.00		2.99	
Total Output	2.993		3.01	
Error %		-.21%		.33%
2. Sugar				
Input	0.0885		0.0885	
Output	0.0896		0.0924	
Error %		1.24%		4.63%
3. First Effect				
Feed	3.00		2.99	
Total Output	2.995		2.985	
Error %		-.15%		-.15%

(Run No. 10L Continued)

	START		FINISH	
4. Second Effect				
Input	1.965		1.910	
Total Output	1.963		1.935	
Error %		-.10%		1.31%
B. Heat Balances (Btu/min.)				
1. First Effect				
Heat Input		1309		1311
Latent Heat	288		260	
Vaporization Heat	1013		1068	
Total Heat Consumed		1301		1328
Heat Losses HL_1		8		-17
Heat Transfer Coefficient (Btu/hr. ft.^{2}°F)	255		199	
2. Second Effect				
Condensing Steam	1013		1068	
Latent Heat	26		38	
Total Heat Input		1039		1106
Total Heat Consumed		985		1085
Heat Losses HL_2		54		21
Heat Transfer Coefficient	450		410	
Inter-Stage Heat Losses HL_{12}		39		34
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	1309		1311	
Feed	186		180	
Total Heat Input Btu/min.	1495		1491	
Heat Outputs:				
First Effect Overhead (Condensate)	162		158	
Second Effect Overhead	1105		1201	
Product	123		94	
Heat Losses	101		38	
Total Heat Output Btu/min.	1491		1491	
Balance Error	-4		0	
C. Calculated Steam Requirements				
Material Balance Steam	1.010*		1.069*	
Latent Heat Steam	.292		.262	
Heat Loss Steam	.101		.038	
Total Steam Required	1.403		1.369	
Actual Steam Flow	1.322		1.322	
Error		6.16%		3.58%

*K_s = 0.500

Run No. 20L

	START		FINISH	
A. Material Balances				
1. Overall	3.070		3.070	
Total Output	3.059		3.005	
Error %		-.36%		-2.12%
2. Sugar				
Input	0.1003		0.0997	
Output	0.1073		0.0997	
Error %		7.00		0.0
3. First Effect				
Feed	3.070		3.070	
Total Output	3.080		3.100	
Error %		.33%		.98%
4. Second Effect				
Input	2.098		1.97	
Total Output	2.077		1.875	
Error %		-1.00%		-4.80%
B. Heat Balance				
1. First Effect				
Heat Input		1346		1548
Latent Heat	301		307	
Vaporization Heat	965		1111	
Total Heat Consumed		1266		1418
Heat Losses HL_1		80		130
Heat Transfer Coefficient	274		286	
2. Second Effect				
Condensing Steam	965		1111	
Latent Heat	48		34	
Total Heat Input		1003		1145
Total Heat Consumed		905		1100
Heat Losses HL_2		98		45
Heat Transfer Coefficient	433		465	
Inter-Stage Heat Losses HL_{12}		50		35
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	1346		1548	
Feed	190		184	
Total Heat Input	1536		1732	
Heat Outputs:				
First Effect Overhead (Condensate)	157		179	

(Run No. 20L Continued)	START		FINISH	
Second Effect Overhead	1020		1236	
Product	150		97	
Heat Losses	228		210	
		1535		1722
Balance Error		-1		-10

C. Calculated Steam Requirements

Material Balance Steam*	.995		1.150	
Latent Heat Steam	.304		.310	
Heat Loss Steam	.231		.212	
Total Steam Required		1.530		1.672
Actual Steam Flow		1.370		1.563
Error		11.70%		6.96%

Run No. 30L

A. Material Balances

1. Overall				
Feed	2.90		2.92	
Total Output	2.935		2.965	
Error %		1.21%		1.54%
2. Sugar				
Input	0.1252		0.0894	
Output	0.1209		0.0920	
Error %		-3.42%		2.91%
3. First Effect				
Feed	2.90		2.92	
Total Output	2.945		2.965	
Error %		1.55%		1.54%
4. Second Effect				
Input	1.925		1.940	
Total Output	1.915		1.940	
Error %		-.52%		0

B. Heat Balances

1. First Effect				
Heat Input		1400		1400
Latent Heat	284		298	
Vaporization Heat	1002		1009	
Total Heat Consumed		1286		1307
Heat Losses HL_1		114		93
Heat Transfer Coefficient	279		288	

* K_s = .505, Large Air Leak @Start.

(Run No. 30L Continued)

	START		FINISH	
2. Second Effect				
Condensing Steam	1002		1009	
Latent Heat	32		34	
Total Heat Input		1034		1043
Total Heat Consumed		909		934
Heat Losses HL_2		125		109
Heat Transfer Coefficient	440		441	
Inter-Stage Heat Losses HL_{12}		33		31
3. Overall Heat Balance				
(Cooling Water Excluded)				
Reference: Water @32°F				
Heat Inputs: Steam (Net)	1400		1400	
Feed	177		166	
Total Heat Input	1577		1566	
Heat Outputs				
First Effect Overhead	162		163	
Second Effect Overhead	1020		1050	
Product	127		127	
Heat Losses	272		233	
	1581		1573	
Balance Error	+4		+7	

C. Calculated Steam Requirements

	START		FINISH	
Material Balance Steam*	.935		.970	
Latent Heat Steam	.286		.300	
Heat Loss Steam	.274		.235	
Total Steam Required		1.496		1.505
Actual Steam Flow		1.417		1.417
Error		5.58%		6.21%

Run No. 40L

A. Material Balances

	START		FINISH	
1. Overall				
Feed	2.92		3.72	
Total Output	2.975		3.837	
Error %		2.57%		3.14%
2. Sugar				
Input	0.0846		0.1098	
Output	0.0926		0.1136	
Error %		9.43%		3.46%

* K_s = .505

(Run No. 40L Continued)

	START		FINISH	
3. First Effect				
Feed	2.92		3.72	
Total Output	2.955		3.79	
Error %		1.20%		1.88%
4. Second Effect				
Input	1.940		2.850	
Total Output	1.960		2.897	
Error %		1.03%		1.65%
B. Heat Balances				
1. First Effect				
Heat Input		1390		1390
Latent Heat	284		360	
Vaporization Heat	998		925	
Total Heat Consumed		1282		1285
Heat Losses HL_1		108		105
Heat Transfer Coefficient	284		279	
2. Second Effect				
Condensing Steam	998		925	
Latent Heat	33		57	
Total Heat Input		1031		982
Total Heat Consumed		935		891
Heat Losses HL_2		96		91
Heat Transfer Coefficient	430		424	
Inter-Stage Heat Losses HL_{12}		33		34
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	1390		1390	
Feed	181		223	
Total Heat Input	1571		1613	
Heat Outputs:				
First Effect Overhead (Condensate)	161		147	
Second Effect Overhead	1050		1000	
Product	129		250	
Heat Losses	237		320	
Total Heat Output	1577		1627	
Balance Error	+6		+14	

C. Calculated Steam Requirements*

	START	FINISH
Material Balance Steam*	1.00	.899
Latent Heat Steam	.286	.362

* K_s = .505

(Run No. 40L Continued)

	START		FINISH	
Heat Loss Steam	.239		.232	
Total Steam Required		1.525		1.493
Actual Steam Flow		1.408		1.408
Error		8.31%		6.05%
Run No. 5FF				
A. Material Balances				
1. Overall				
Feed	2.98		2.98	
Total Output	2.985		3.005	
Error %		.165%		.84%
2. Sugar				
Input	0.0915		0.1290	
Output	0.0976		0.1347	
Error %		6.66%		4.42%
3. First Effect				
Feed	2.98		2.98	
Total Output	3.01		3.005	
Error %		1.01%		.84%
4. Second Effect				
Input	2.03		2.34	
Total Output	2.005		2.34	
Error %		-1.23%		0
B. Heat Balances				
1. First Effect				
Heat Input		1181		843
Latent Heat	278		250	
Vaporization Heat	966		660	
Total Heat Consumed		1244		910
Heat Losses HL_1		-63		-67
Heat Transfer Coefficient	205		177	
2. Second Effect				
Condensing Steam	996		660	
Latent Heat	33		30	
Total Heat Input		1029		690
Total Heat Consumed		936		644
Heat Losses HL_2		93		46
Heat Transfer Coefficient	502		402	
Inter-Stage Heat Losses HL_{12}		28		26

(Run No. 5FF Continued)

3. Overall Heat Balance

	START		FINISH	
Heat Inputs: Steam (Net)	1181		843	
Feed	182		188	
Total Heat Input	1363		1031	
Heat Outputs:				
First Effect Overhead (Condensate)	150		98	
Second Effect Overhead	1049		721	
Product	133		208	
Heat Losses	58		5	
		1390		1032
Balance Error		+27		+1

C.

	START	FINISH
Overall $K_s = \dfrac{S}{W_o(1-C_o/C_p)}$	.711	.775
Computer K_s	.710	.710

Run No. 6FF

A. Material Balances

1. Overall

	START		FINISH	
Feed	2.99		2.99	
Total Output	3.072		3.066	
Error %		2.74%		2.54%

2. Sugar

	START		FINISH	
Input	0.0924		0.1043	
Output	0.0965		0.1078	
Error %		4.44%		3.45%

3. First Effect

	START		FINISH	
Feed	2.98		2.98	
Total Output	2.982		2.996	
Error %		.07%		.54%

4. Second Effect

	START		FINISH	
Input	2.20		2.370	
Total Output	2.29		2.390	
Error %		4.31%		.85%

B. Heat Balance

1. First Effect

	START		FINISH	
Heat Input		1035		925

(Run No. 6FF Continued)

	START		FINISH	
Latent Heat	265		260	
Vaporization Heat	760		657	
Total Heat Consumed		1025		917
Heat Losses HL_1		10		+8
Heat Transfer Coefficient	208		192	
2. Second Effect				
Condensing Steam	760		657	
Latent Heat	29		28	
Total Heat Input		789		685
Total Heat Consumed		765		633
Heat Losses HL_2		34		52
Heat Transfer Coefficient	445		401	
Inter-Stage Heat Losses HL_{12}		29		31
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	1035		925	
Feed	179		182	
Total Heat Input	1214		1107	
Heat Outputs:				
First Effect Overhead (Condensate)	115		84	
Second Effect Overhead	857		710	
Product	188		226	
Heat Losses	73		91	
Total Heat Output		1233		1111
Balance Error		+19		+4

C. Calculated Steam Requirements

	START		FINISH	
Material Balance Steam*	.767		.650	
Latent Heat Steam	.265		.260	
Heat Loss Steam	.073		.091	
Total Steam Required		1.105		1.001
Actual Steam Flow		1.035		.921
Error		+2.90%		+8.56%

Run No. 7FF

A. Material Balances

	START		FINISH	
1. Overall				
Feed	2.99		3.00	
Total Output	3.04		3.050	
Error %		1.67%		1.64%

* $K_s = .505$

(Run No. 7FF Continued)

	START		FINISH	
2. Sugar				
Input	0.1048		0.1035	
Output	0.1082		0.1079	
Error %		3.25%		4.25%
3. First Effect				
Feed	2.99		3.00	
Total Output	3.047		3.045	
Error %		1.91%		1.50%
4. Second Effect				
Input	2.375		1.985	
Total Output	2.368		1.990	
Error %		-.30%		.25%
B. Heat Balance				
1. First Effect				
Heat Input		920		1300
Latent Heat	260		285	
Vaporization Heat	665		1047	
Total Heat Consumed		925		1332
Heat Losses HL_1		-5		-32
Heat Transfer Coefficient (Btu/hr. ft.^{2}°F)	201		240	
2. Second Effect				
Condensing Steam	665		1047	
Latent Heat	29		26	
Total Heat Input		694		1073
Total Heat Consumed		621		1005
Heat Losses HL_2		73		68
Heat Transfer Coefficient	405		479	
Inter-Stage Heat Losses HL_{12}		29		38
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	920		1300	
Feed	182		183	
Total Heat Input	1102		1483	
Heat Outputs				
First Effect Overhead (Condensate)	99		164	
Second Effect Overhead	696		1128	
Product	216		122	
Heat Losses	97		74	
Total Heat Output	1108		1488	
Balance Error	+6		+5	

(Run No. 7FF Continued)

	START		FINISH	
C. Calculated Steam Requirements				
Material Balance Steam*	.650		1.025	
Latent Heat Steam	.259		.288	
Heat Loss Steam	.097		.075	
Total Steam Required		1.006		1.388
Actual Steam Flow		.915		1.316
Error		9.94%		5.48%
Run No. 8FF				
A. Material Balances				
1. Overall				
Feed	2.98		2.98	
Total Output	3.061		3.066	
Error %		2.72%		2.88%
2. Sugar				
Input	0.1029		0.1268	
Output	0.1098		0.1370	
Error %		6.70%		8.05%
3. First Effect				
Feed	2.98		2.98	
Total Output	3.04		3.016	
Error %		2.02%		1.21%
4. Second Effect				
Input	1.98		2.10	
Total Output	2.001		2.150	
Error %		1.06%		2.38%
B. Heat Balance				
1. First Effect				
Heat Input		1300		1140
Latent Heat	283		259	
Vaporization Heat	1042		905	
Total Heat Consumed		1325		1164
Heat Losses HL_1		-25		-24
Heat Transfer Coefficient	246		216	

* K_s = .505

(Run No. 8FF Continued)

	START		FINISH	
2. Second Effect				
Condensing Steam	1042		905	
Latent Heat	26		27	
Total Heat Input		1068		932
Total Heat Consumed		1004		875
Heat Losses HL_2		64		57
Heat Transfer Coefficient	492		490	
Inter-Stage Heat Losses HL_{12}		38		32
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	1300		1140	
Feed	194		191	
Total Heat Input	1494		1331	
Heat Outputs:				
First Effect Overhead (Condensate)	165		138	
Second Effect Overhead	1129		980	
Product	125		158	
Heat Losses	77		65	
Total Heat Output	1496		1341	
Balance Error	+2		+10	

C. Calculated Steam Requirements

	START		FINISH	
Material Balance Steam*	1.022		.900	
Latent Heat Steam	.286		.260	
Heat Loss Steam	.078		.065	
Total Steam Required		1.386		1.225
Actual Steam Flow		1.315		1.149
Error		5.40%		6.62%

Run No. 9FF

A. Material Balances

	START		FINISH	
1. Overall				
Feed	2.98		3.85	
Total Output	3.040		3.895	
Error %		2.02%		1.17%
2. Sugar				
Input	0.1281		0.1654	
Output	0.1350		0.1702	
Error %		4.60%		2.90%

* K_s = .500

(Run No. 9FF Continued)

	START		FINISH	
3. First Effect				
Feed	2.98		3.85	
Total Output	2.995		3.85	
Error %		.50%		0
4. Second Effect				
Input	2.085		2.730	
Total Output	2.130		2.775	
Error %		2.16%		1.65%
B. Heat Balance				
1. First Effect				
Heat Input		1140		1450
Latent Heat	259		358	
Vaporization Heat	899		1101	
Total Heat Consumed		1158		1459
Heat Losses HL_1		210		258
Heat Transfer Coefficient		-18		-9
2. Second Effect				
Condensing Steam	899		1101	
Latent Heat	27		60	
Total Heat Input		926		1161
Total Heat Consumed		875		1117
Heat Losses HL_2		51		44
Heat Transfer Coefficient	487		414	
Inter-Stage Heat Losses HL_{12}		31		33
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	1140		1450	
Feed	194		246	
Total Heat Input	1334		1696	
Heat Outputs:				
First Effect Overhead (Condensate)	132		176	
Second Effect Overhead	980		1250	
Product	156		204	
Heat Losses	64		68	
Total Heat Output	1332		1698	
Balance Error	-2		+2	
C. Calculated Steam Requirements				
Material Balance Steam*	.892		1.123	
Latent Heat Steam	.261		.362	

* $K_s = .500$

(Run No. 9FF Continued)

	START		FINISH	
Heat Loss Steam	.065		.069	
Total Steam Required		1.218		1.554
Actual Steam Flow		1.149		1.470
Error		6.05%		5.71%

Run No. 10FF

A. Material Balances

	START		FINISH	
1. Overall				
Feed	2.98		2.98	
Total Output	2.98		3.005	
Error %		0.0		0.84%
2. Sugar				
Input	0.0805		0.0829	
Output	0.0824		0.0842	
Error %		2.36%		1.57%
3. First Effect				
Feed	2.98		2.98	
Total Output	3.01		3.00	
Error %		1.01%		0.67%
4. Second Effect				
Input	1.89		1.950	
Total Output	1.86		1.955	
Error %		-1.59%		0.26%

B. Heat Balance

	START		FINISH	
1. First Effect				
Heat Input		1370		1280
Latent Heat	280		274	
Vaporization Heat	1102		1038	
Total Heat Consumed		1382		1312
Heat Losses HL_1		-12		-32
Heat Transfer Coefficient	269		236	
2. Second Effect				
Condensing Steam	1102		1038	
Latent Heat	28		27	
Total Heat Input		1130		1065
Total Heat Consumed		1075		990
Heat Losses HL_2		55		75
Heat Transfer Coefficient	475		490	
Inter-Stage Heat Losses HL_{12}		36		33

(Run No. 10FF Continued)

	START	FINISH
3. Overall Heat Balance		
Heat Inputs: Steam (Net)	1370	1280
Feed	191	188
Total Heat Input	1561	1468
Heat Outputs:		
First Effect Overhead (Condensate)	177	162
Second Effect Overhead	1208	1120
Product	98	120
Heat Losses	79	76
Total Heat Output	1562	1478
Balance Error	+1	+10

C. Calculated Steam Requirements

	START		FINISH	
Material Balance Steam*	1.106		1.012	
Latent Heat Steam	.283		.277	
Heat Loss Steam	.080		.077	
Total Steam Required		1.469		1.366
Actual Steam Flow		1.388		1.292
Error		5.81%		5.72%

Run No. 11FFB

A. Material Balances

	START		FINISH	
1. Overall				
Feed	2.97		2.97	
Total Output	3.058		3.045	
Error %		2.62%		2.52%
2. Sugar				
Input	0.1021		0.1173	
Output	0.1116		0.1190	
Error %		9.30%		1.45%
3. First Effect				
Feed	2.97		2.97	
Total Output	3.00		3.02	
Error %		1.01%		1.68%
4. Second Effect				
Input	2.00		2.05	
Total Output	2.058		2.075	
Error %		2.90%		1.22%

* K_s = .500

(Run No. 11FFB Continued)

	START		FINISH	
B. Heat Balance				
1. First Effect				
Heat Input		1251		1210
Latent Heat	268		273	
Vaporization Heat	987		957	
Total Heat Consumed		1255		1230
Heat Losses HL_1		-4		-20
Heat Transfer Coefficient	254		224	
2. Second Effect				
Condensing Steam	987		957	
Latent Heat	28		29	
Total Heat Input		1015		986
Total Heat Consumed		962		925
Heat Losses HL_2		53		61
Heat Transfer Coefficient	522		505	
Inter-Stage Heat Losses HL_{12}		32		33
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	1251		1210	
Feed	190		184	
Total Heat Input	1441		1394	
Heat Outputs:				
First Effect Overhead (Condensate)	154		149	
Second Effect Overhead	1080		1038	
Product	136		143	
Heat Losses	81		74	
Total Heat Output	1451		1404	
Balance Error	+10		+10	
C. Calculated Steam Requirements				
Material Balance Steam*	.981		.915	
Latent Heat Steam	.269		.275	
Heat Loss Steam	.082		.075	
Total Steam Required		1.332		1.265
Actual Steam Flow		1.260		1.210
Error		5.71%		4.55%

* K_s = .500

Run No. 12FFB

	START		FINISH	
A. Material Balances				
1. Overall				
Feed	2.97		2.97	
Total Output	3.003		2.993	
Error %		1.10%		0.79%
2. Sugar				
Input	0.1003		0.1129	
Output	0.1056		0.1140	
Error %		5.3%		0.975%
3. First Effect				
Feed	2.97		2.97	
Total Output	2.995		2.988	
Error %		.841%		.605%
4. Second Effect				
Input	2.05		2.09	
Total Output	2.058		2.095	
Error %		.39%		.24
B. Heat Balance				
1. First Effect				
Heat Input		1180		1118
Latent Heat	285		276	
Vaporization Heat	932		886	
Total Heat Consumed		1217		1162
Heat Losses HL_1		-37		-44
Heat Transfer Coefficient (Btu/hr.ft.^{2}°F)	211		212	
2. Second Effect				
Condensing Steam	932		886	
Latent Heat	29		31	
Total Heat Input		961		917
Total Heat Consumed		892		855
Heat Losses HL_2		69		62
Heat Transfer Coefficient	455		432	
Inter-Stage Heat Losses HL_{12}		33		31
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	1180		1118	
Feed	175		184	
Total Heat Input		1355		1302

(Run No. 12FFB Continued)

	START		FINISH	
Heat Outputs:				
First Effect Overhead (Condensate)	146		139	
Second Effect Overhead	1001		960	
Product	146		155	
Heat Losses	65		49	
Total Heat Output	1358		1303	
Balance Error	+3		+1	

C. Calculated Steam Requirements

	START		FINISH	
Material Balance Steam*	.927		.868	
Latent Heat Steam	.288		.278	
Heat Loss Steam	.066		.049	
Total Steam Required		1.281		1.195
Actual Steam Flow		1.194		1.125
Error		7.27%		6.23%

Run No. 15FFB

A. Material Balances

	START		FINISH	
1. Overall				
Feed	3.00		3.870	
Total Output	2.809		3.910	
Error %		-6.35%		1.03%
2. Sugar				
Input	0.1293		0.1650	
Output	0.1148		0.1730	
Error %		-11.2%		4.85%
3. First Effect				
Feed	3.00		3.87	
Total Output	2.959		3.87	
Error %		-1.37%		0
4. Second Effect				
Input	2.16		2.81	
Total Output	2.01		2.85	
Error %		-6.94%		1.42%

* K_s = .500

(Run No. 15FFB Continued)

B. Heat Balance

	START		FINISH	
1. First Effect				
Heat Input		993		1343
Latent Heat	255		344	
Vaporization Heat	790		1042	
Total Heat Consumed		1045		1386
Heat Losses HL$_1$		-52		-43
Heat Transfer Coefficient	224		263	
2. Second Effect				
Condensing Steam	790		1042	
Latent Heat	30		56	
Total Heat Input		820		1098
Total Heat Consumed		754		1055
Heat Losses HL$_2$		66		43
Heat Transfer Coefficient	413		477	
Inter-Stage Heat Losses HL$_{12}$		30		34
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	993		1343	
Feed	204		263	
Total Heat Input		1197		1606
Heat Outputs				
First Effect Overhead (Condensate)	122		166	
Second Effect Overhead	845		1183	
Product	157*		224	
Heat Losses	44		34	
Total Heat Output	1168		1607	
Balance Error	-29		+1	

C. Calculated Steam Requirements

	START		FINISH	
Material Balance Steam**	.790		1.073	
Latent Heat Steam	.258		.348	
Heat Loss Steam	.045		.034	
Total Steam Required		1.093		1.455
Actual Steam Flow		1.001		1.360
Error %		9.19%		7.00%

* Product Flow Too Low

** K_s = .500

Run No. 16FFB

	START		FINISH	
A. Material Balances				
1. Overall				
Feed	2.97		2.97	
Total Output	3.01		3.022	
Error %		1.35%		1.75%
2. Sugar				
Input	0.0810		0.0809	
Output	0.0796		0.0864	
Error %		-1.73%		6.80%
3. First Effect				
Feed	2.97		2.97	
Total Output	3.01		3.00	
Error %		1.35%		1.01%
4. Second Effect				
Input	2.000		1.850	
Total Output	2.000		1.872	
Error %		0		1.19%
B. Heat Balance				
1. First Effect				
Heat Input		1211		1400
Latent Heat	264		270	
Vaporization Heat	998		1132	
Total Heat Consumed		1262		1402
Heat Losses HL_1		-51		-2
Heat Transfer Coefficient		134		242
2. Second Effect				
Condensing Steam	998		1132	
Latent Heat	34		30	
Total Heat Input		1032		1162
Total Heat Consumed		1036		1178
Heat Losses HL_2		-4		-16
Heat Transfer Coefficient	470		474	
Inter-Stage Heat Losses HL_{12}		28		35
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	1211		1400	
Feed	190		195	
Total Heat Input		1401		1595
Heat Outputs:				
First Effect Overhead (Condensate)	154		180	
Second Effect Overhead	1160		1319	

(Run No. 16FFB Continued)

	START		FINISH	
Product	118		85	
Heat Losses	-27		17	
Total Heat Output		1405		1601
Balance Error		+4		+6

C. Calculated Steam Requirements

	START		FINISH	
Material Balance Steam*	.998		1.154	
Latent Heat Steam	.272		.278	
Heat Loss Steam	-.028		.018	
Total Steam Required		1.250		1.450
Actual Steam Flow		1.242		1.500
Error %		0.64%		3.33%

Run No. 17FB

A. Material Balances

	START		FINISH	
1. Overall				
Feed	2.99		3.00	
Total Output	2.99		3.03	
Error %		0.0		1.00%
2. Sugar				
Input	0.1354		0.0924	
Output	0.1336		0.0886	
Error %		-1.33%		-4.11%
3. First Effect				
Feed	2.99		3.00	
Total Output	2.98		3.045	
Error %		-.33%		1.50%
4. Second Effect				
Input	2.18		1.955	
Total Output	2.19		1.940	
Error %		.46%		-.77%

B. Heat Balance

	START		FINISH	
1. First Effect				
Heat Input		1041		1332
Latent Heat	254		273	
Vaporization Heat	791		1074	
Total Heat Consumed		1045		1347
Heat Losses HL_1		-4		-15
Heat Transfer Coefficient	171		208	

* K_s = .500

(Run No. 17FB Continued)

	START		FINISH	
2. Second Effect				
Condensing Steam	791		1074	
Latent Heat	24		27	
Total Heat Input		815		1101
Total Heat Consumed		819		1075
Heat Losses HL_2		-4		26
Heat Transfer Coefficient	429		491	
Inter-Stage Heat Losses HL_{12}		35		35
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	1041		1332	
Feed	195		195	
Total Heat Input		1236		1527
Heat Outputs:				
First Effect Overhead (Condensate)	120		169	
Second Effect Overhead	930		1206	
Product	169		108	
Heat Losses	27		46	
Total Heat Output		1246		1529
Balance Error		+10		+2
C. Calculated Steam Requirements				
Material Balance Steam*	.796		1.047	
Latent Heat Steam	.256		.278	
Heat Loss Steam	.027		.047	
Total Steam Required		1.079		1.367
Actual Steam Flow		1.050		1.357
Error %		2.75%		.73%

Run No. 18FB

A. Material Balances

	START		FINISH	
1. Overall				
Feed	3.00		3.48	
Total Output	3.015		3.48 (Balanced to Get O_1)	
Error %		.50%		0.0
2. Sugar				
Input	0.0924		0.1044	
Output	0.0870		0.1119	
Error %		-5.85%		7.16%

* K_s = .500

(Run No. 18FB Continued)

	START		FINISH	
3. First Effect				
Feed	3.00		3.48	
Total Output	2.99		3.445	
Error %		-.33%		-1.00%
4. Second Effect				
Input	1.900		2.250	
Total Output	1.925		2.285	
Error %		1.32%		1.56%
B. Heat Balance				
1. First Effect				
Heat Input		1332		1579
Latent Heat	270		334	
Vaporization Heat	1075		1175	
Total Heat Consumed		1345		1509
Heat Losses HL_1		-13		70
Heat Transfer Coefficient	208		241	
2. Second Effect				
Condensing Steam	1075		1175	
Latent Heat	27		43	
Total Heat Input		1102		1218
Total Heat Consumed		1090		1236
Heat Losses HL_2		12		-18
Heat Transfer Coefficient	508		490	
Inter-Stage Heat Losses HL_{12}		32		36
3. Overall Heat Balance				
Heat Inputs: Steam (Net)	1332		1579	
Feed	195		222	
Total Heat Input		1527		1799
Total Heat Input				
First Effect Overhead (Condensate)	168		189	
Second Effect Overhead	1221		1387	
Product	104		131	
Heat Losses	31		88	
Total Heat Output		1524		1795
Balance Error		-3		-4
C. Calculated Steam Requirements				
Material Balance Steam*	1.052		1.249	
Latent Heat Steam	.274		.342	

* K_s = .500

(Run No. 18FB Continued)

	START		FINISH	
Heat Loss Steam	.032		.909	
Total Steam Required		1.358		1.681
Actual Steam Flow		1.355		1.610
Error %		+0.41%		+4.41%

APPENDIX G

MISCELLANEOUS EXPERIMENTAL RUNS

The experimental runs reported in this appendix are each in some manner different from the runs presented in the main body of this thesis. The first was subjected to a disturbance entirely different from those of other runs, the second was made using the original feedforward control scheme (which proved too general to be accurate), and the remainder were somewhat less than successful runs due to improper controller settings within the feedforward unit or on the Foxboro instrumentation.

These runs were included as a part of this thesis because they were the forerunners of the more successful runs, and they do contribute towards a better understanding of problems which may be encountered in the design of a suitable control system for a complicated unit such as a multiple effect evaporator. The following section summarizes the description of each experimental run, its observed transient behavior and the comments which are pertinent to each run.

The transient responses of some of the major process variables are presented for each experimental run in the figures following the summary.

Summary of Miscellaneous Runs

1. Run No. 10L

Disturbance: Condenser pressure stepped from 4.41 to 3.00 Psia.

Product Concentration Response:

- A rapid and significant increase resulted in C_p.

Comments:

- Liquid levels were relatively unaffected.
- The condenser condensate removal system was satisfactory even as the condenser pressure approached the vacuum supply pressure
- Noise in condenser flow did increase.
- Second effect circulation pump cavitated continuously.
- First effect bottoms flow was disturbed by the cavitation in the circulating line.

2. Run No. 5FF

Disturbance: Feed composition stepped from 3.07 to 4.31% sugar.

Product Concentration Response:

- Final study state product concentration was approximately 1% sugar below the set point.

Comments:

- Steam Economy, as originally defined, attempted to represent heat losses, latent heat requirements, and heat required to concentrate the feed, by a constant K_s. The total steam requirement was calculated using a value of $K_s = 0.710$ in the equation
$$S = W_o K_s (1 - C_o / C_{psp}).$$
- It is evident that this defination of K_s is too general and each of the three heat requirements mentioned above should be computed independently.

3. Run No. 6FF

Disturbance: Feed concentration stepped from 3.10% to 3.43% sugar.

Product Concentration Response:

- Permanent offset in C_p : 0.20% sugar too low.

Comments:

- K_s was set at 0.505.
- Heat loss bias appeared to be too low at the final steady state. It was increased following 6FF.

4. Run No. 7FF

Disturbance: Product concentration set point stepped from 6.10% to 10.0% sugar.

Product Concentration Response:

- Permanent offset in C_p : 0.89% sugar too high.

Comments:
- Heat loss bias now appeared too high.
- K_s was set at 0.505.

The product concentration response in runs numbered 6FF and 7FF indicated that K_s might also be set slightly too high. It was reset at 0.500 and the heat loss bias was reset to balance the steam flow at the final steady state rate for Run No. 7FF.

5. Run No. 11FFB

Disturbance: Feed concentration stepped from 3.44% to 3.90% sugar. Foxboro concentration controller settings: 50% P.B. and 5 min./repeat.

Product Concentration Response:

- Slightly oscillatory around set point after two hours (caused by steam flow oscillations).

Comments:
- Feedback control action of the concentration controller is too great.
- Response is too oscillatory even for small feed concentration disturbances.
- Decrease % P.B. and increase reset time.

6. Run No. 13FFB

Disturbance: Feed flow stepped from 2.98 to 3.87 lbs./min. Foxboro concentration controller settings: 130% P.B. and 10 min./repeat.

Product Concentration Response:

- Unstable.

Comments:
- Since the evaporator is more sensitive to steam flow changes that it is to feed flow changes, it is reasonable to assume that the above settings would also produce instability following product concentration set point changes.

7. Run No. 14FFB

Disturbance: Feed flow stepped from 3.84 to 3.00 lbs./min. Foxboro concentration controller settings: 400% P.B. and 2 min./repeat.

Product Concentration Response:

- Unstable.

Comments: - The feedback control settings were calculated using a procedure outlined on Page 91 of Buckley's book, "Techniques of Process Control" (3). This procedure is for the control of dead time systems only. The severe instability which resulted demonstates that these settings are not satisfactory when there are also capacity elements in the process.

On the basis of the observations made during Runs 13FFB and 14FFB, it was concluded that only a very little feedback action could be added to the feedforward circuit if severe oscillation of the product concentration was to be avoided following a feed flow disturbance or a concentration set point change. The control action was cut to 400% proportional band and 30 minutes per repeat before further runs were made.

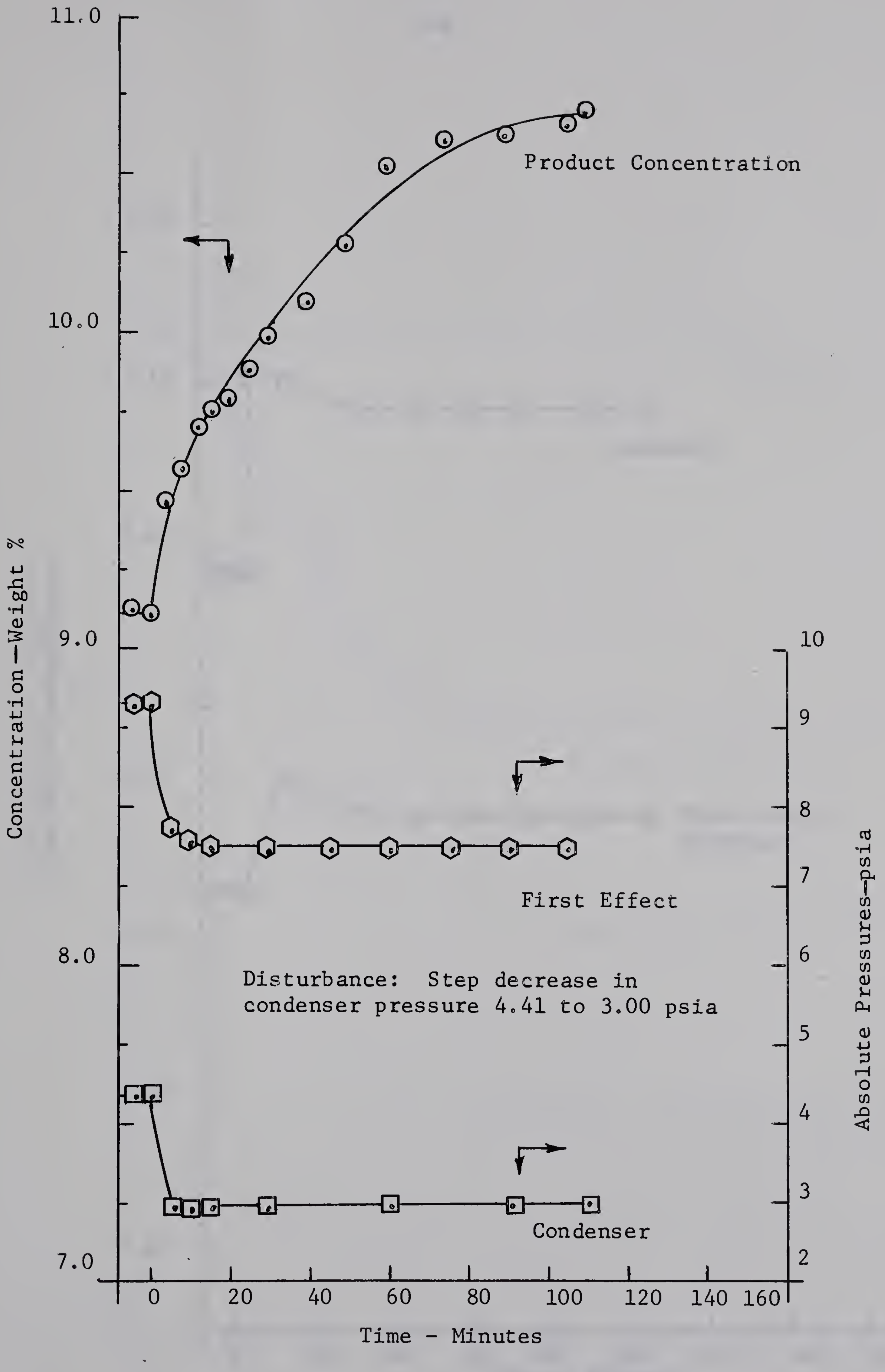


Figure G-1 CONCENTRATION AND PRESSURE TRANSIENT RESPONSES FOR RUN 10L

[illegible]

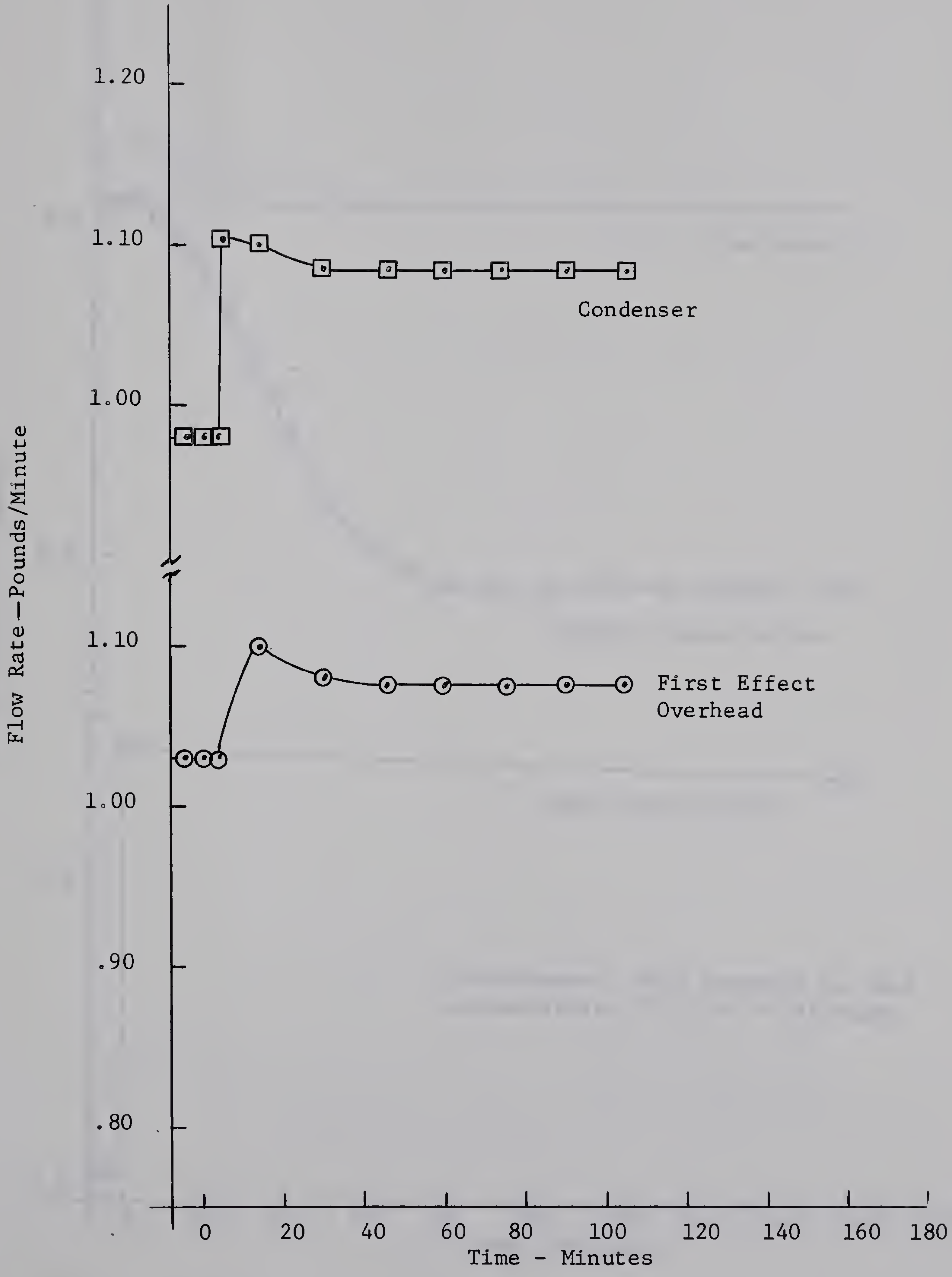
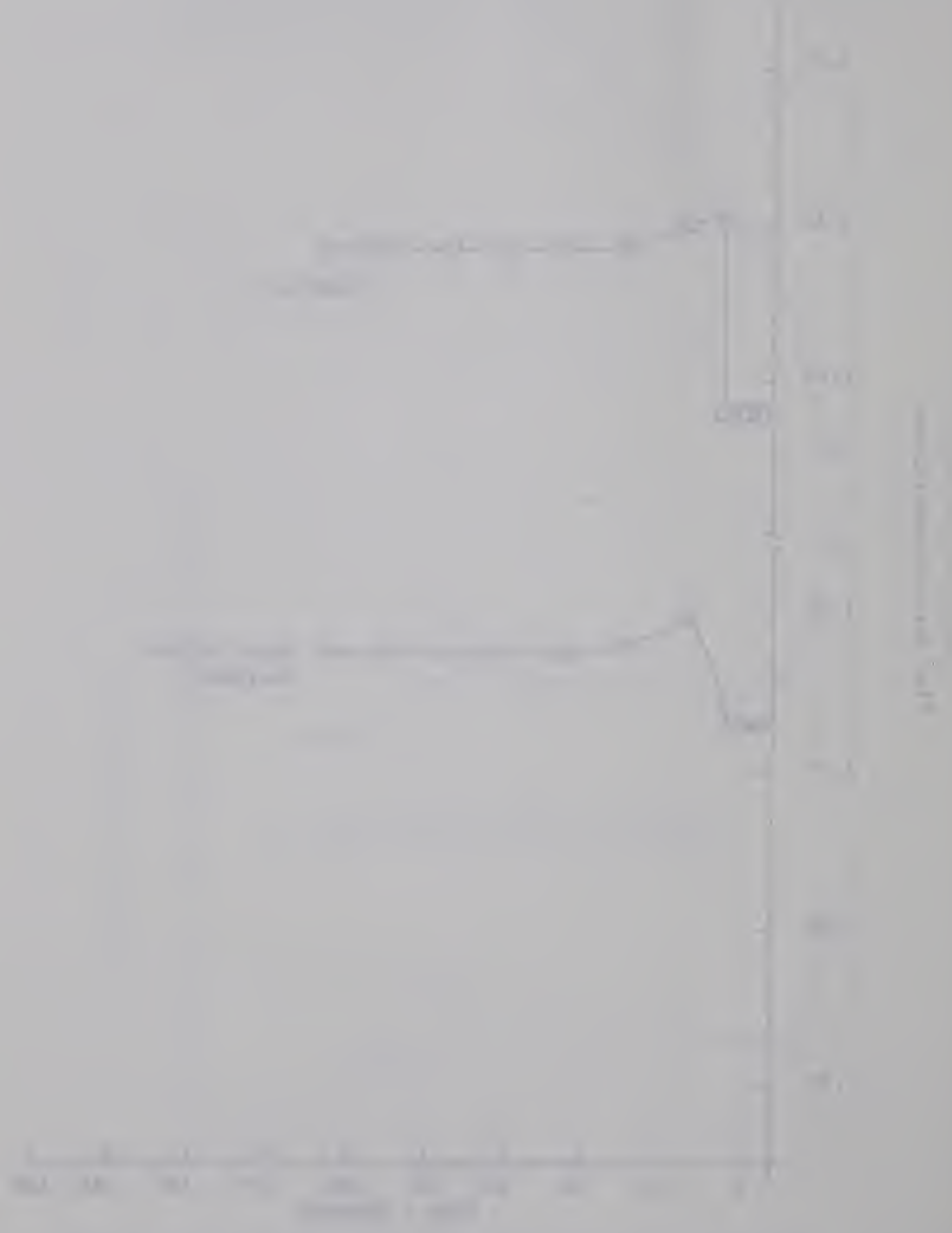


Figure G-2 TRANSIENT RESPONSE OF CONDENSATE FLOWS FOR RUN 10L



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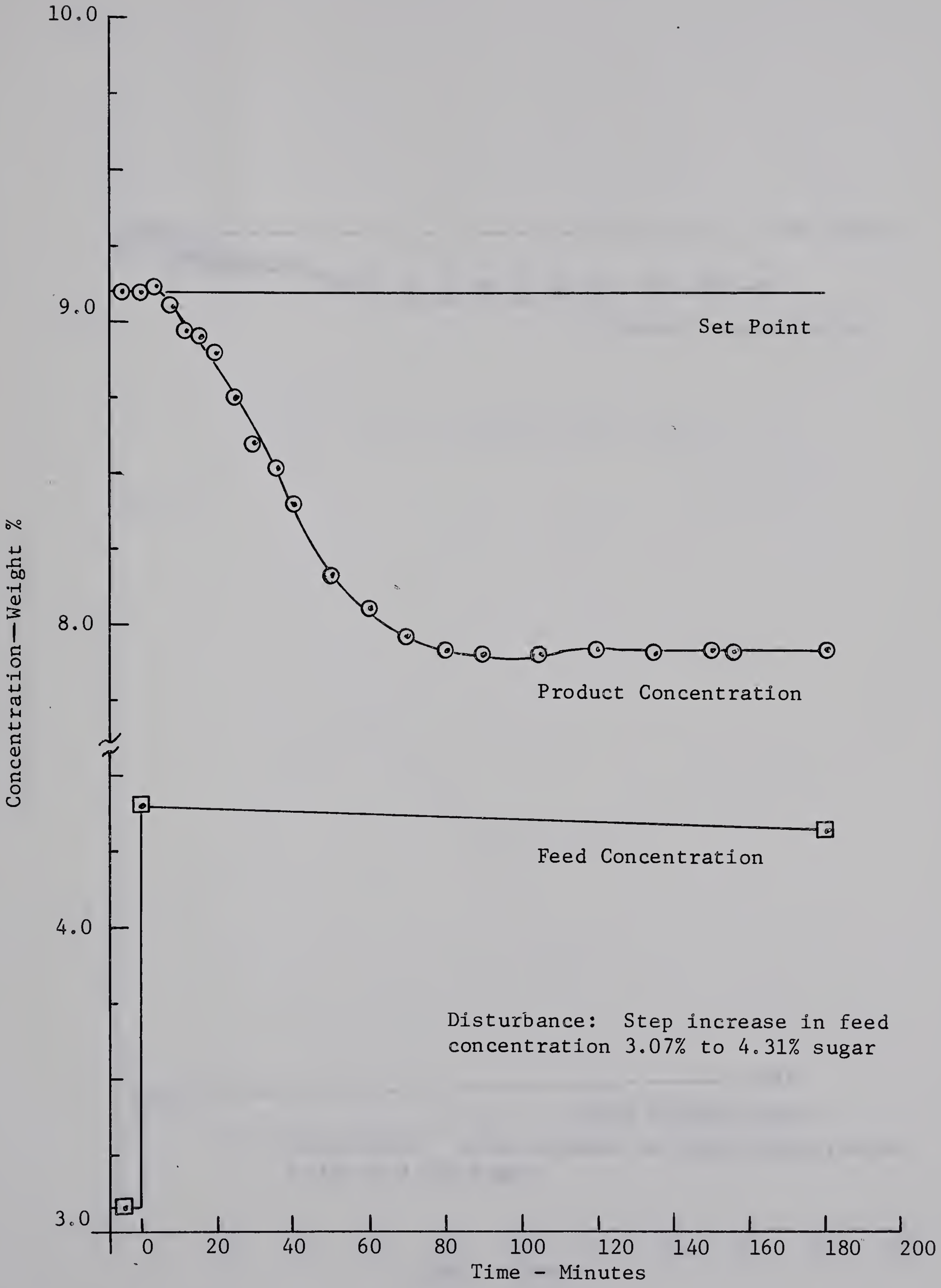


Figure G-3 TRANSIENT CONCENTRATION RESPONSE FOR RUN 5FF

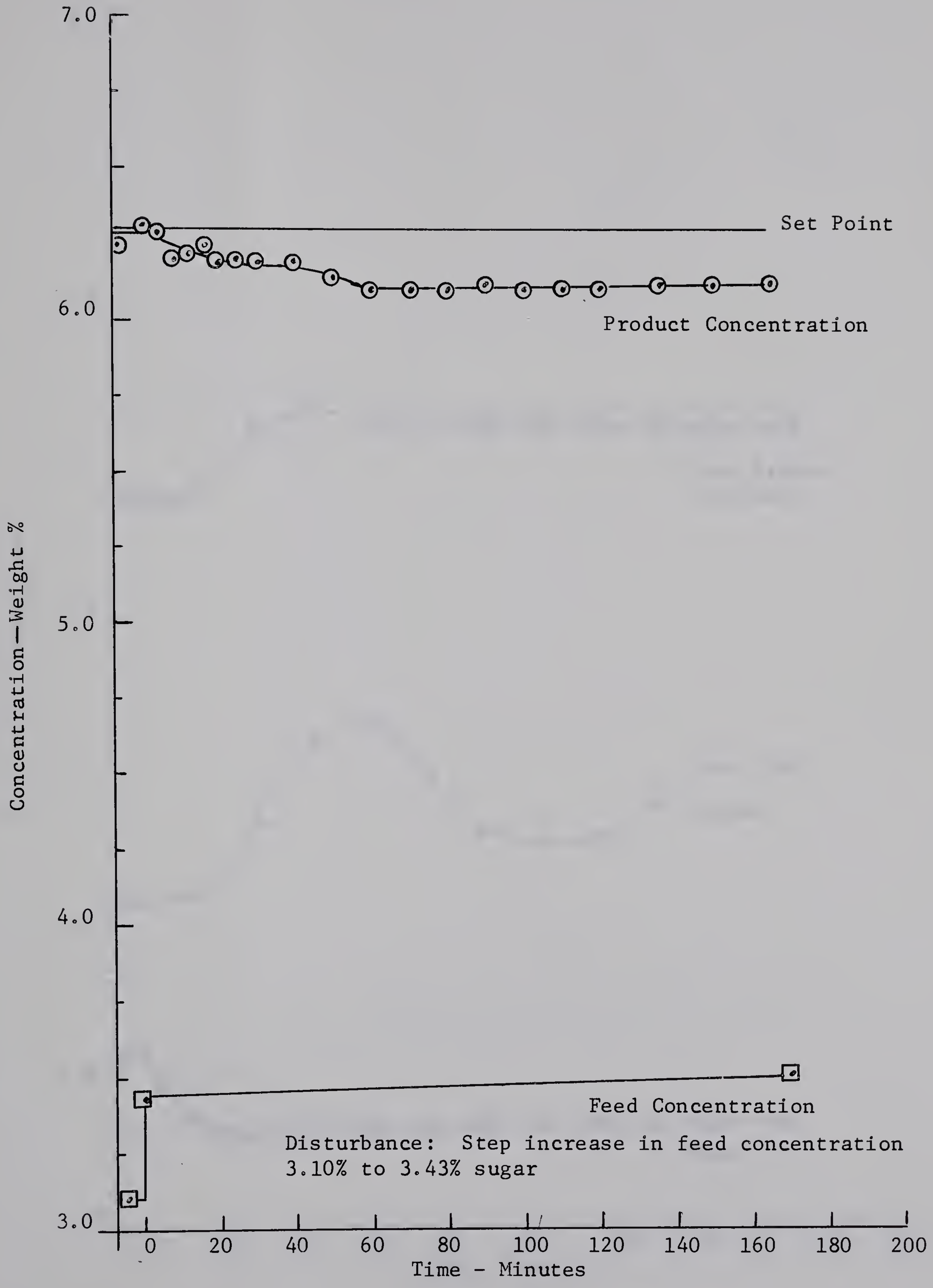


Figure G-4 TRANSIENT CONCENTRATION RESPONSE FOR RUN 6FF

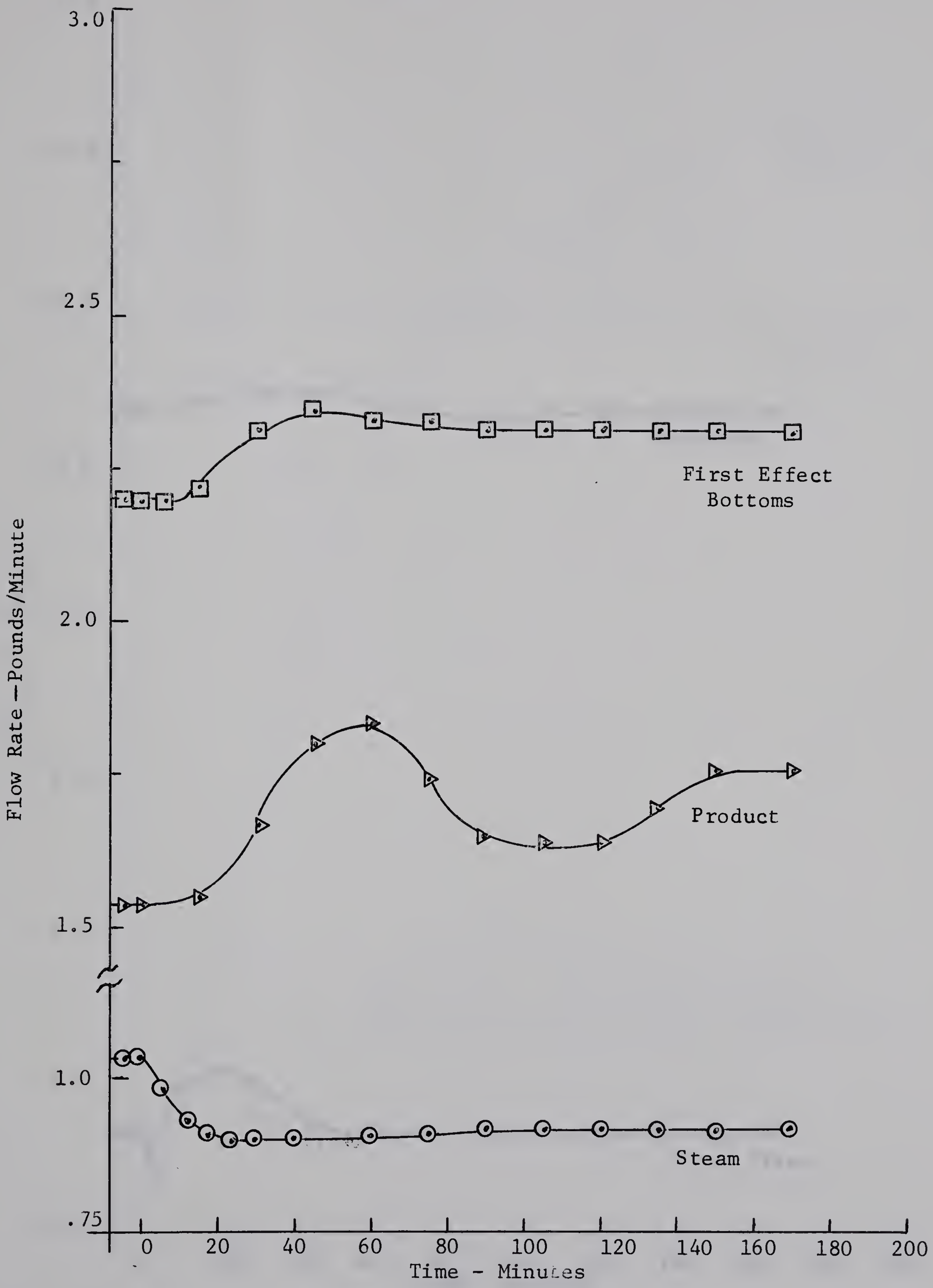


Figure G-5 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN 6FF

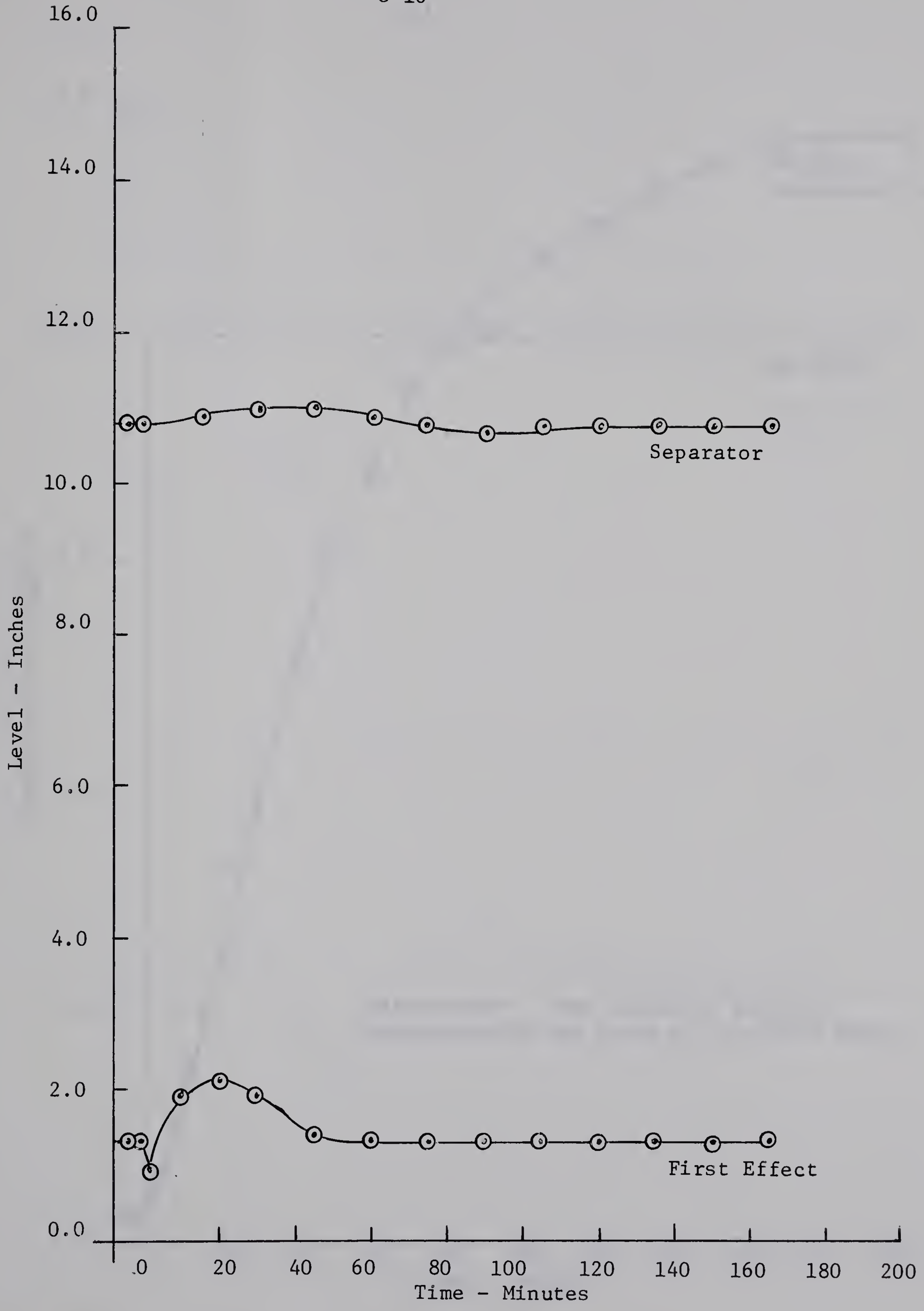


Figure G-6 TRANSIENT RESPONSE OF LIQUID LEVELS FOR RUN 6FF

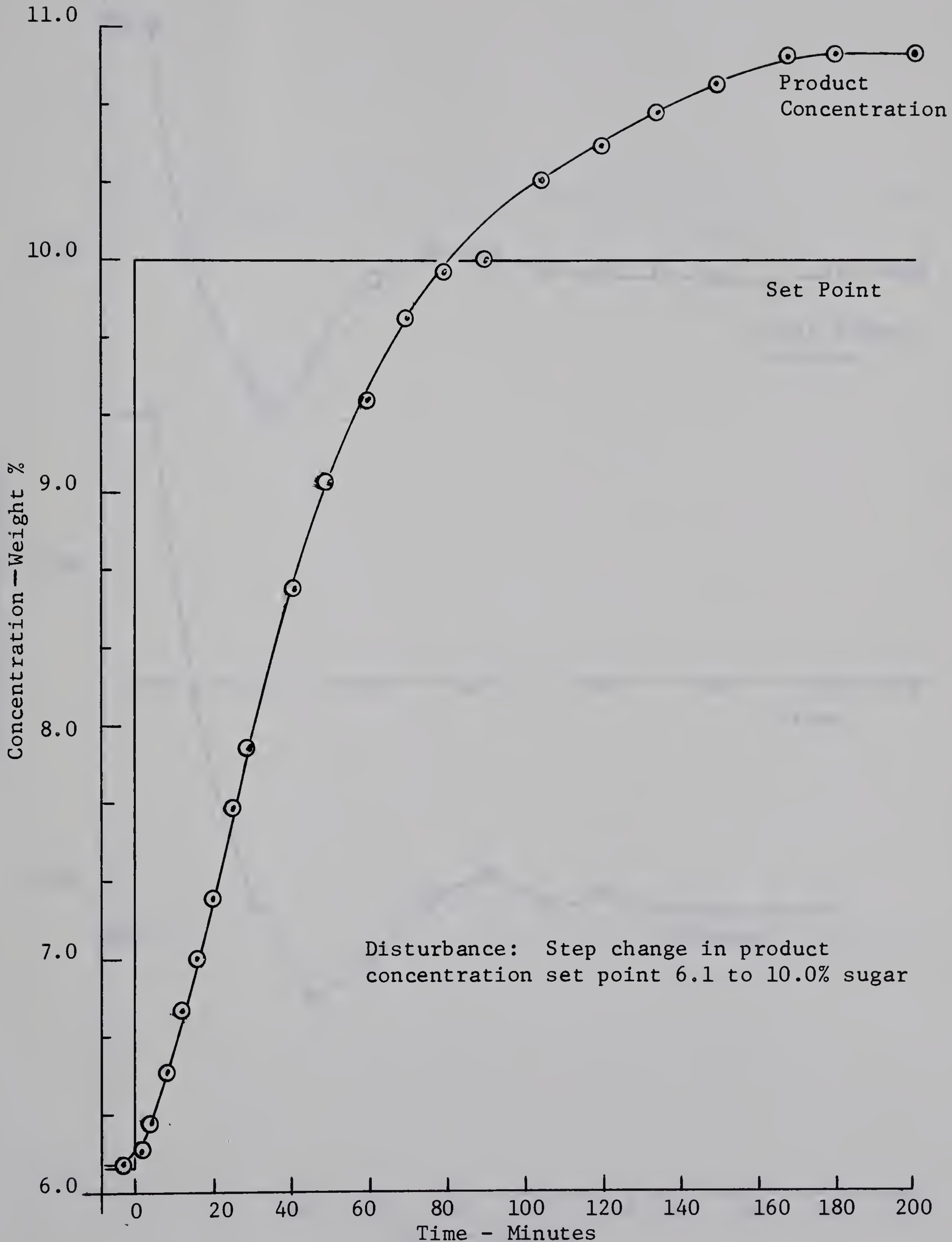
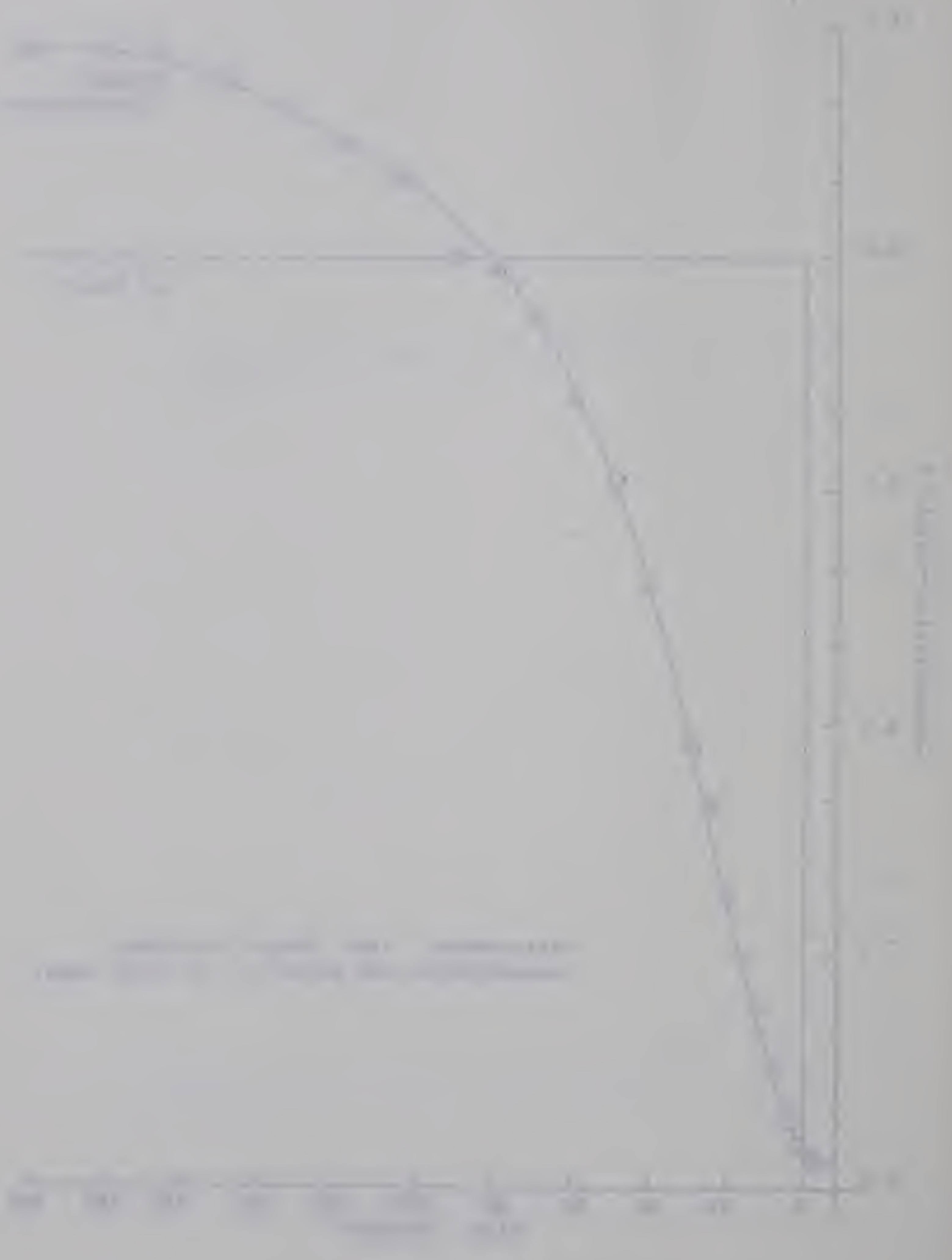


Figure G-7 TRANSIENT CONCENTRATION RESPONSE FOR RUN 7FF



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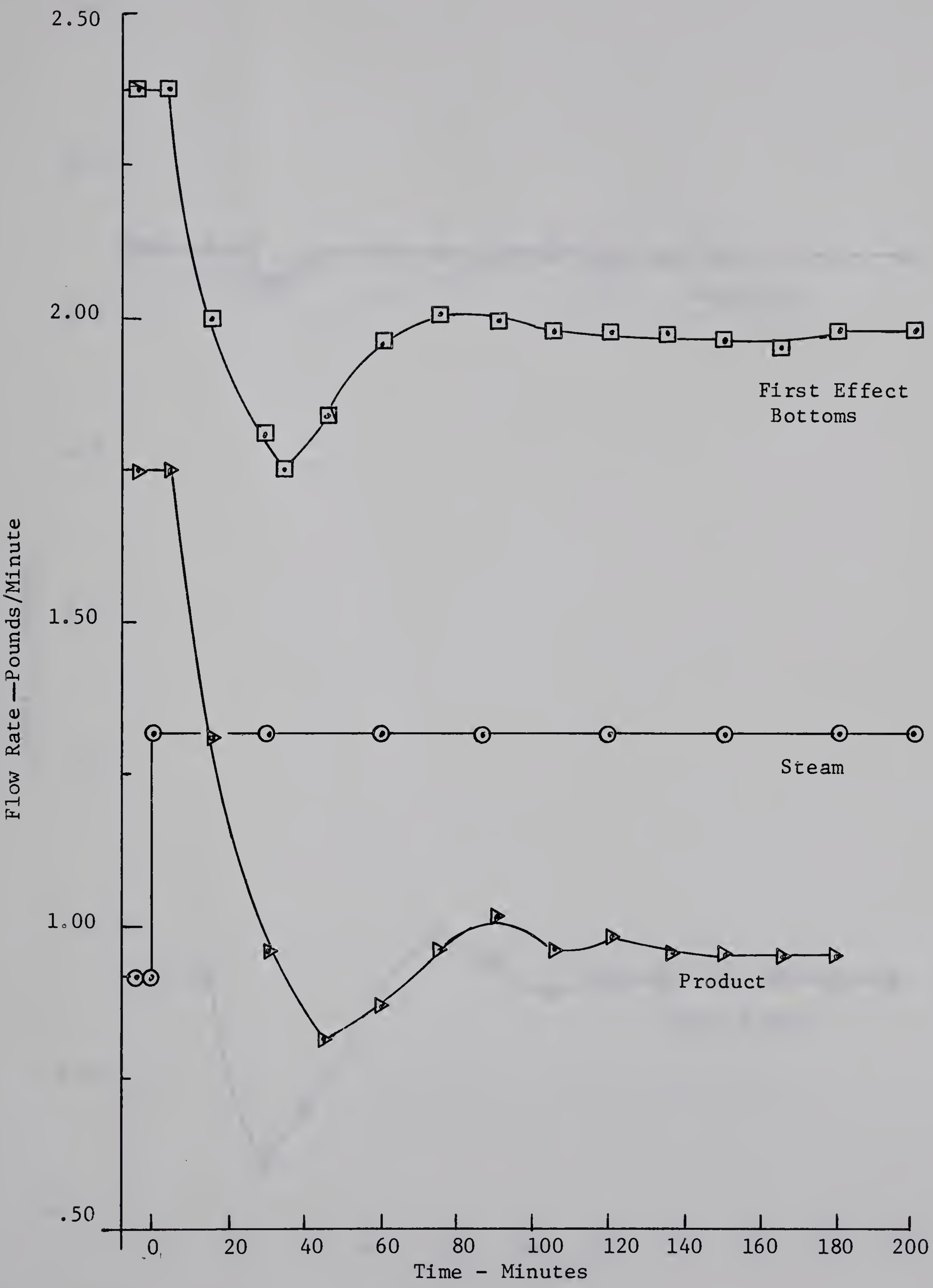


Figure G-8 TRANSIENT RESPONSE OF SYSTEM FLOWS FOR RUN 7FF

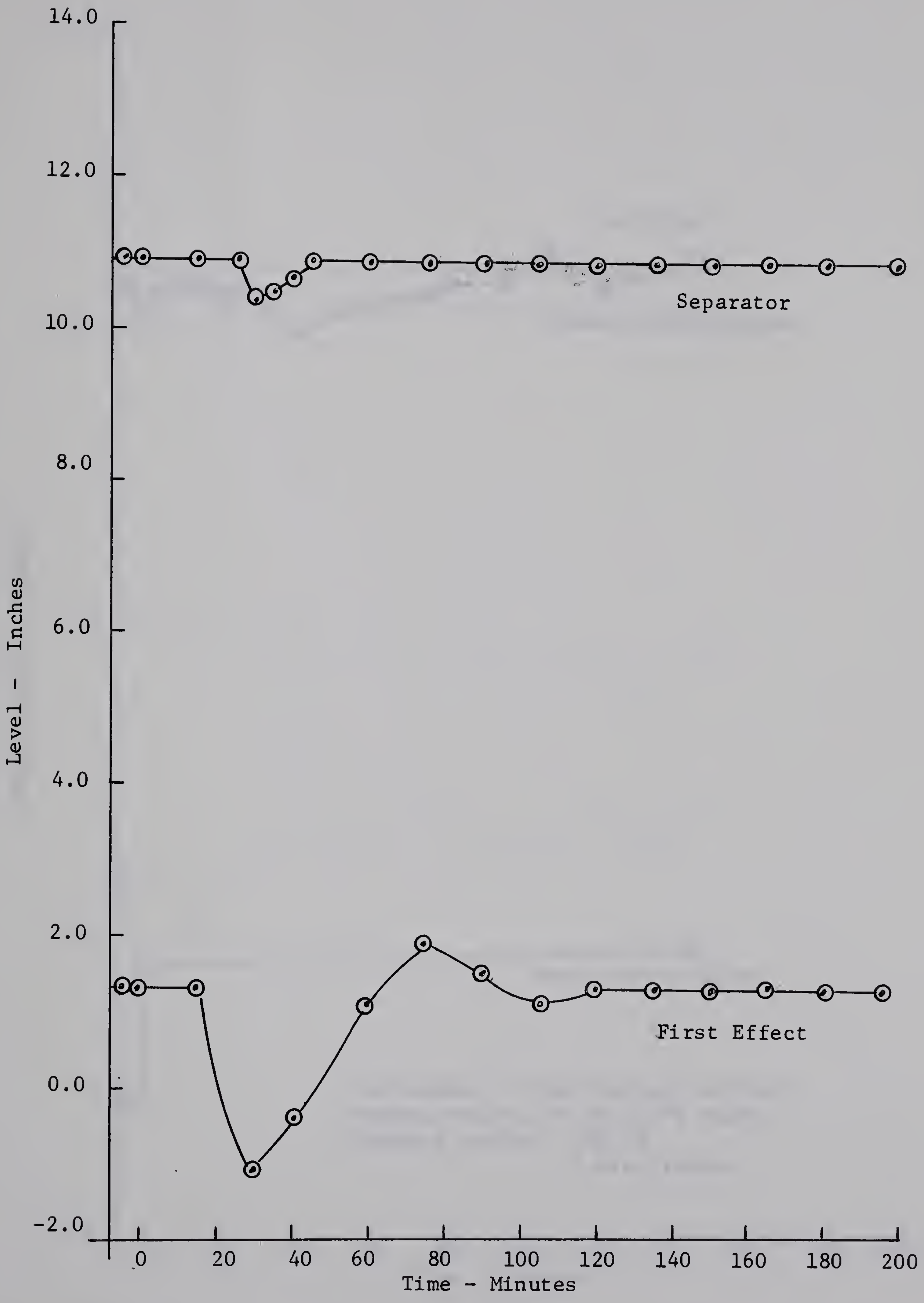
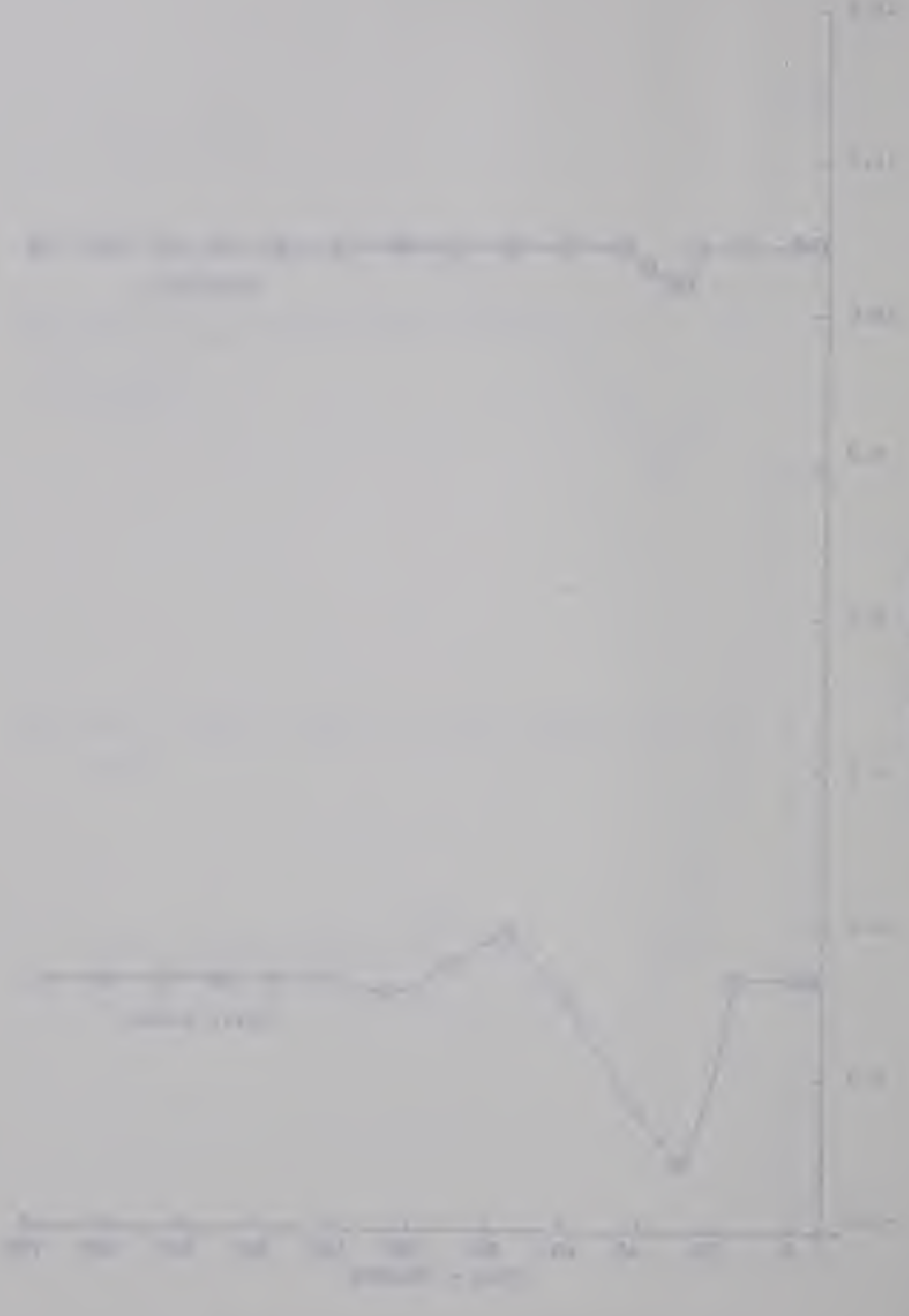
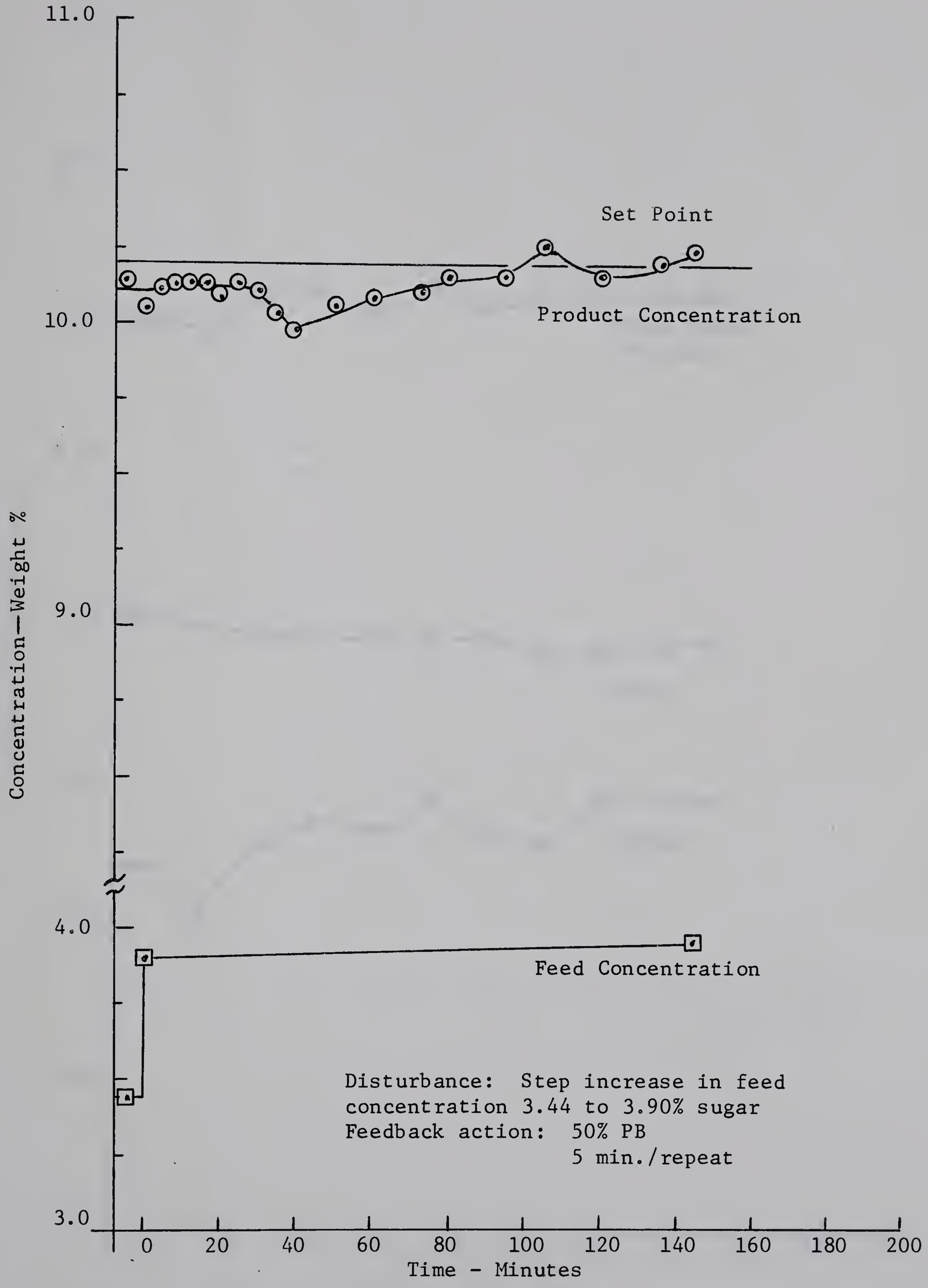


Figure G-9 TRANSIENT RESPONSE OF LIQUID LEVELS FOR RUN 7FF

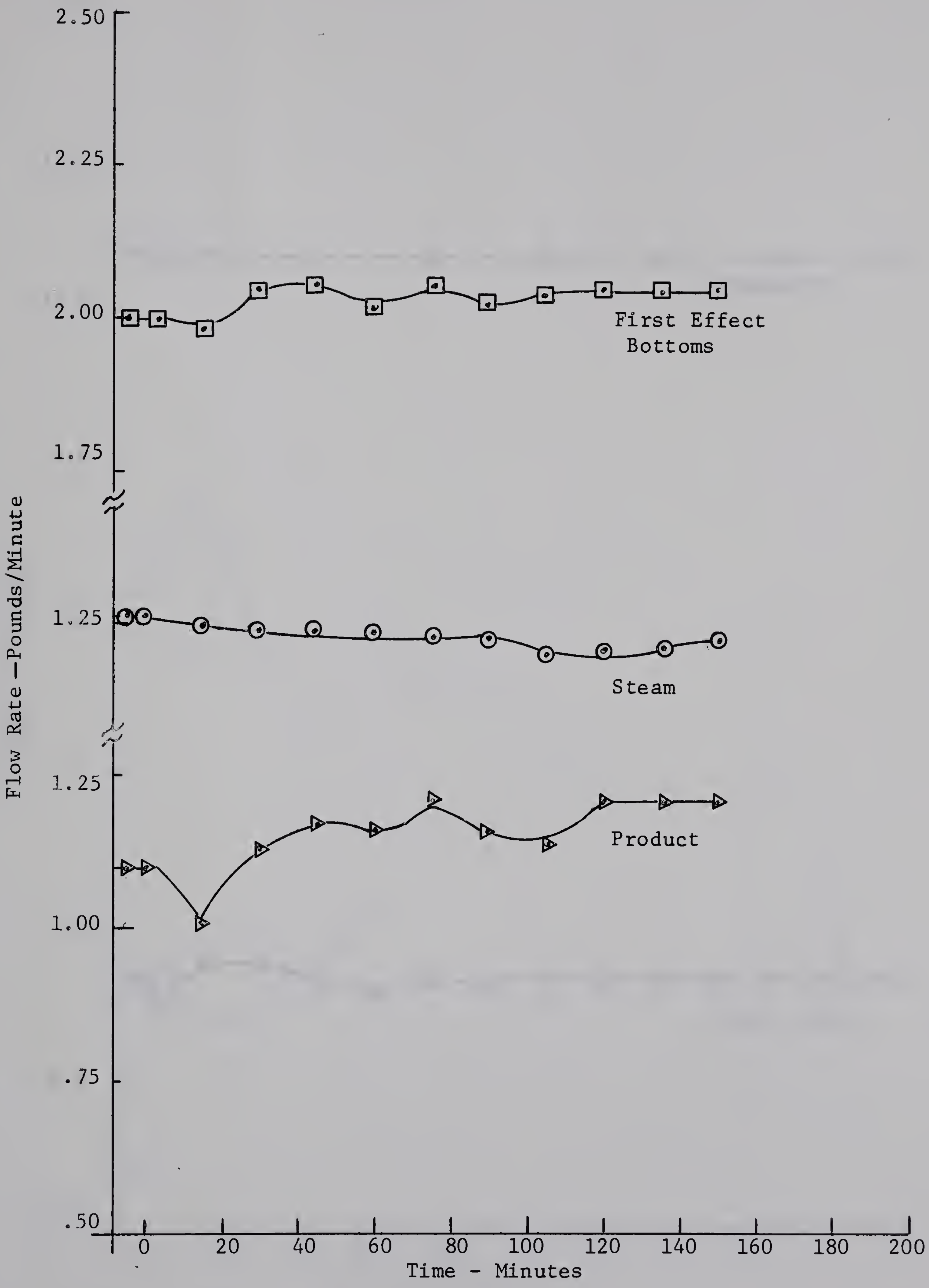


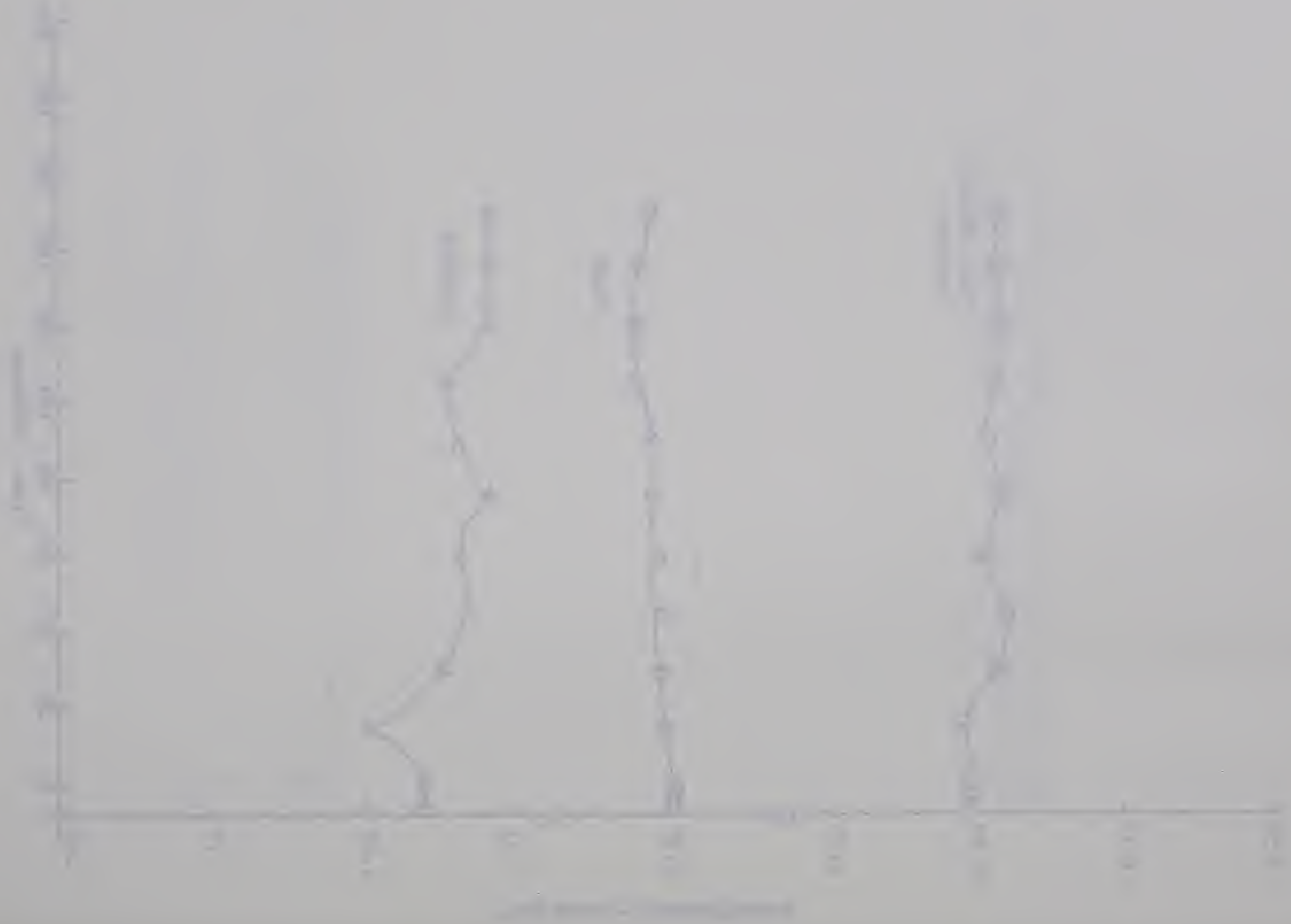
[illegible]



Disturbance: Step increase in feed concentration 3.44 to 3.90% sugar
Feedback action: 50% PB
5 min./repeat

Figure G-10 TRANSIENT CONCENTRATION RESPONSE FOR RUN 11FFB


2.50
2.25
2.00
1.75
1.25
1.25
1.00
.75
.50
Flow Rate —Pounds/Minute
0
20
40
60
80
100
120
140
160
180
200
Time - Minutes
First Effect
Bottoms
Steam
Product



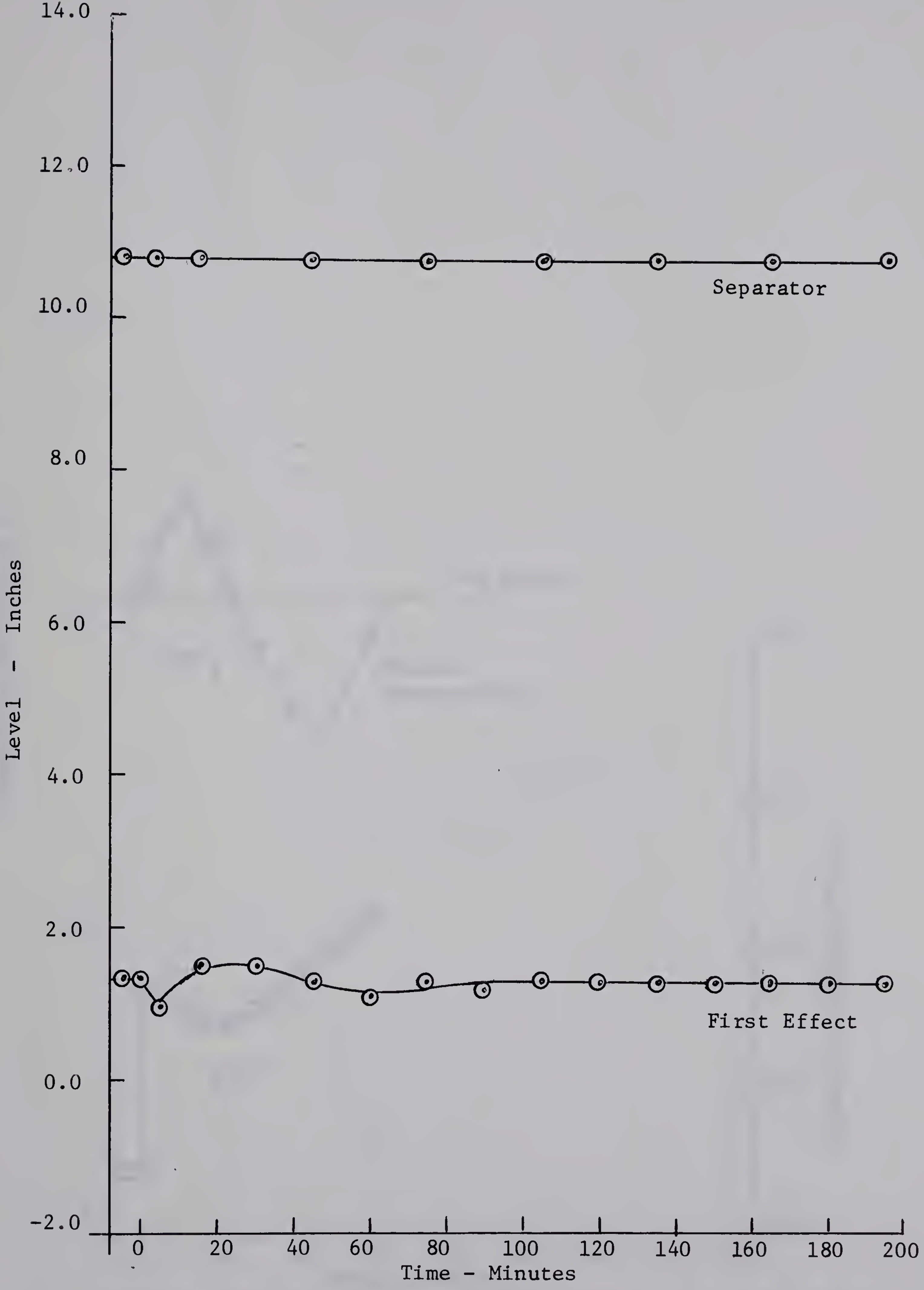
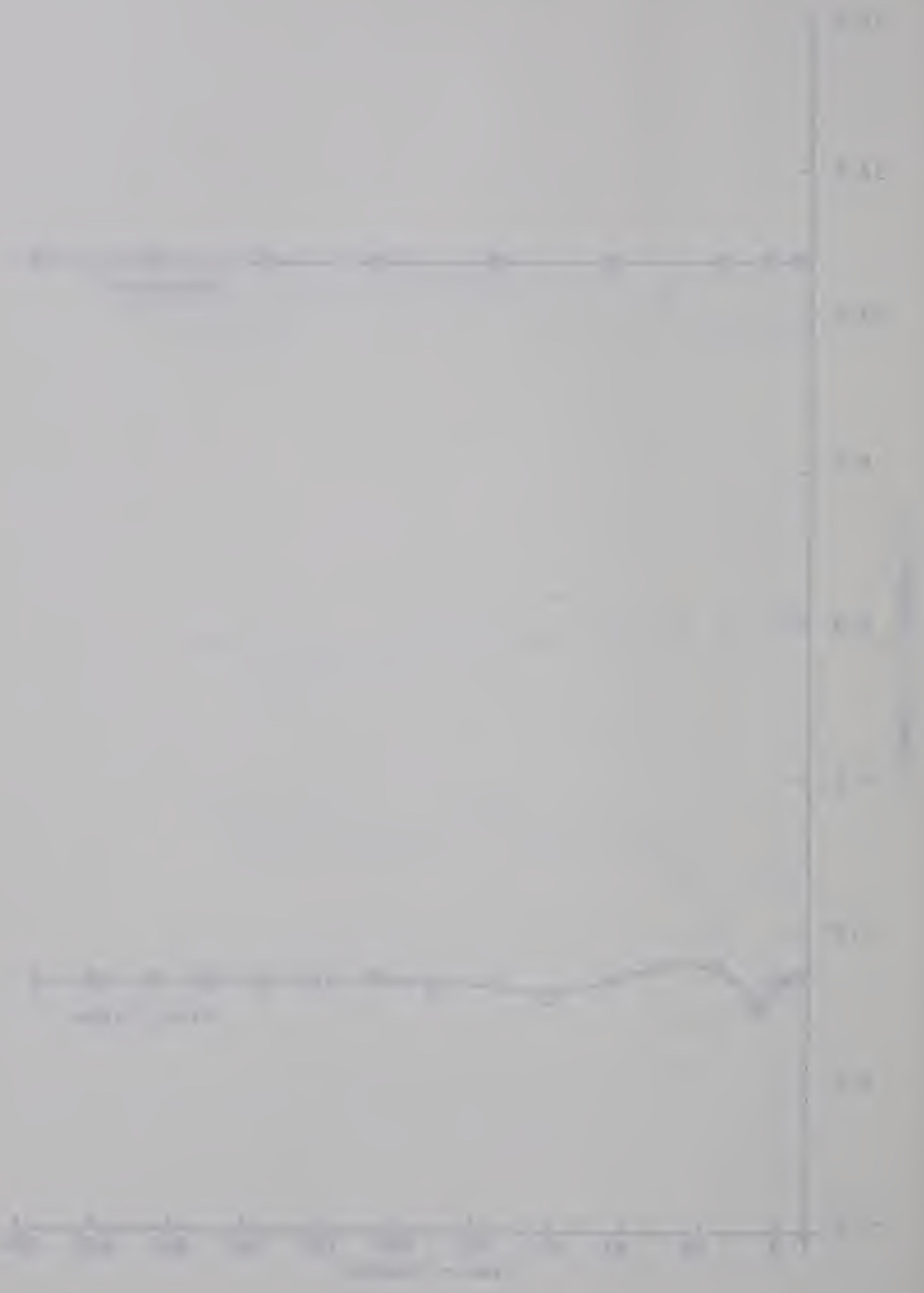
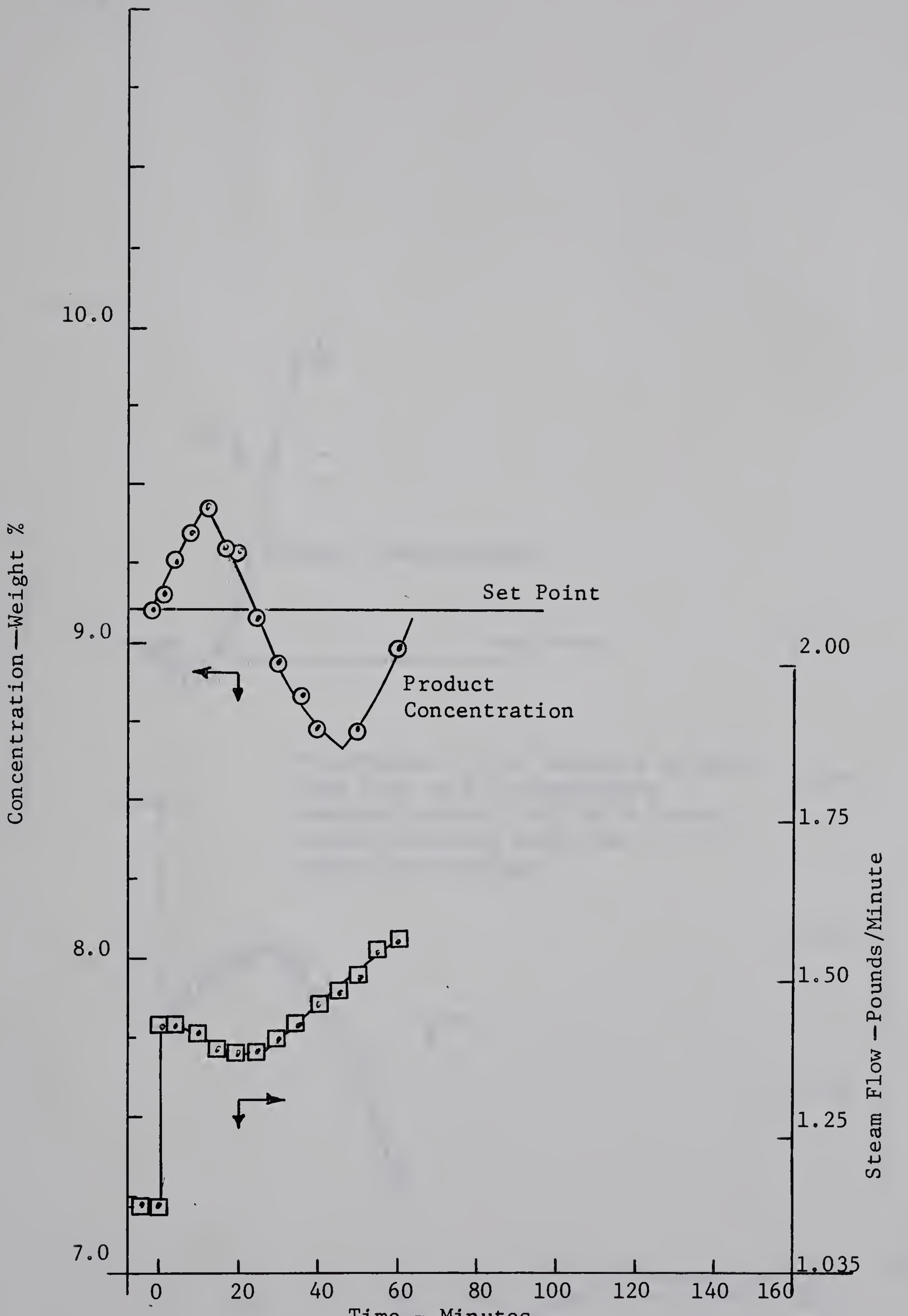


Figure G-12 TRANSIENT RESPONSE OF SYSTEM LEVELS FOR RUN 11FFB



[illegible]



Disturbance: Step increase in feed flow 2.98 to 3.87 lbs./min. Feedback action 130% PB&10 min./repeat (Response using 225% PB & 15 min./repeat only slightly better).

Figure G-13 CONCENTRATION AND STEAM TRANSIENTS FOR RUN NO. 13FFB

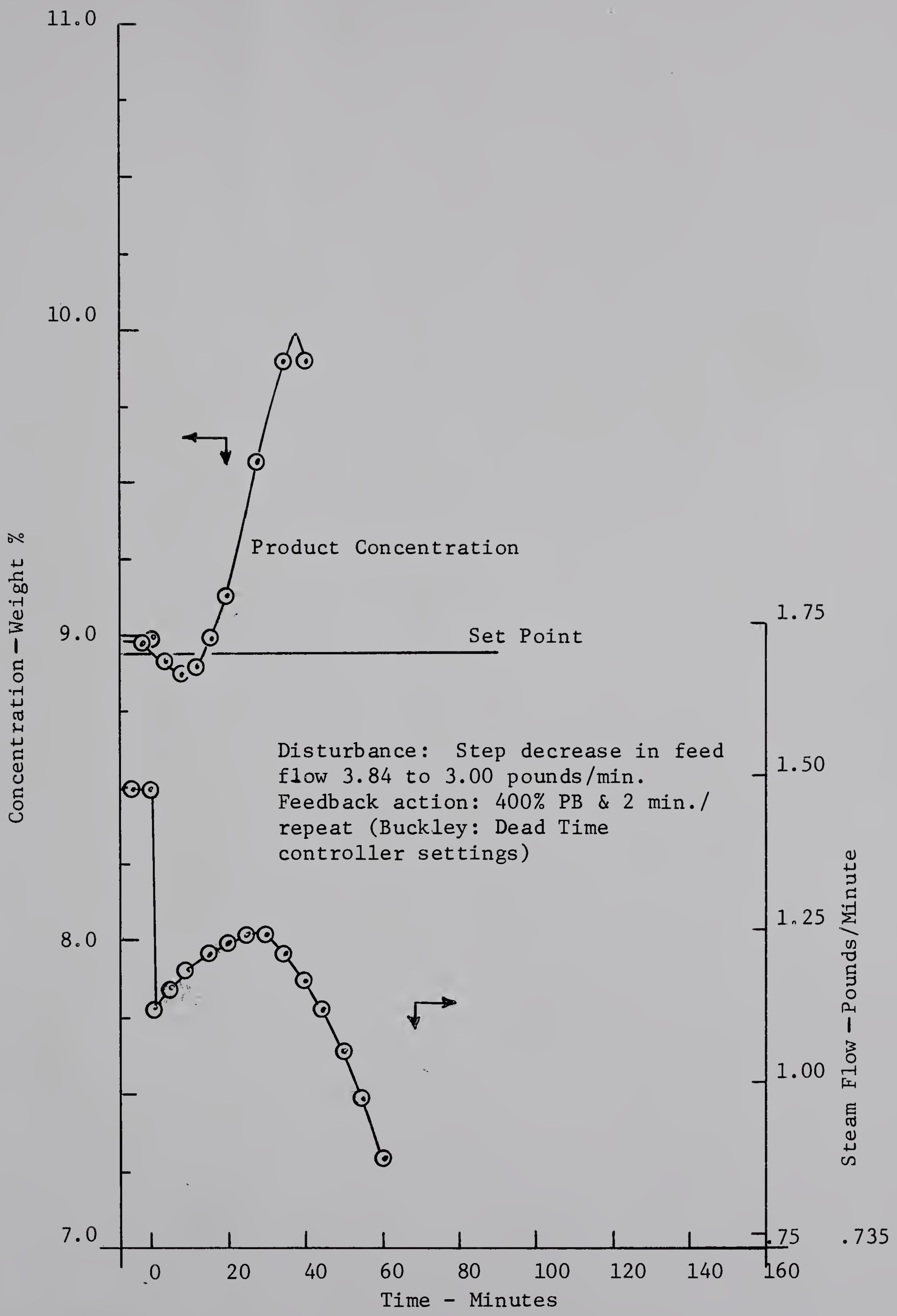


Figure G-14 CONCENTRATION AND STEAM TRANSIENT RESPONSES FOR RUN 14FFB

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